Hierarchical Procedure for Plantwide Control System Synthesis

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A hierarchical procedure for synthesizing an optimal plantwide control system is proposed for an existing continuous chemical process. Alternative plantwide control systems are synthesized and are compared based on economics. The cost associated with dynamic controllability is quantified by the controllability index ν introduced by Zheng and Mahajanam (1999). The procedure is illustrated on a simple reactor-separator-recycle system. Although the procedure is discussed for an existing plant, a simple modification of the approach can be used to determine the optimum surge capacities of a process, including reflux drums and column sumps—a problem of increasing interest.

Introduction

Plantwide control refers to the control of an entire plant, consisting of many interconnected unit operations. While extensive research has been conducted on the control of individual unit operations, relatively few attempts have been made on plantwide control, which has been well reviewed by Skogestad and Larsson (1998). With increasing process integration through material and energy recycles, and reduction in surge capacities, a unit operation-based approach for plantwide control is becoming inadequate. A systems viewpoint should be taken.

The pioneer work on plantwide control was carried out by Buckley (1964), who introduced the concept of "dynamic process control" in the 1960s. He proposed decomposing the problem based on time-scale differences (such as product quality controllers are much faster than material balance controllers). Findeisen et al. (1980) first introduced the concept of "feedback optimizing control." Arkun (1978) used this concept for the selection of controlled variables. Morari et al. (1980) presented a unified formulation for the problem of synthesizing control structures for chemical processes. Based on extensive dynamic simulation studies of a reactor-separator-recycle system, Price and Georgakis (1993) proposed a tiered design framework and derived several guidelines for the plantwide control system design problem. Based on their extensive work on case studies (Tyreus and Luyben, 1993; Luyben and Luyben, 1995; Yi and Luyben, 1995; Luyben et al., 1996; Lyman et al., 1996), Luyben et al. (1997) recently proposed a nine-step procedure for plantwide control system design. The procedure synthesizes only one control structure and does not consider any alternative control structure. Wolff and Skogestad (1994) offered a number of qualitative guidelines for finding control strategies for integrated processes, while Morud and Skogestad (1993) discussed the effects of various interconnections (such as material recycle, energy recycle, and so on) on overall process behavior. Skogestad and Postlethwaite (1996) gave an excellent discussion on some of the steps involved on the design of a plantwide control system and discussed the concept of self-optimizing control. Skogestad and co-workers (Halvorsen and Skogestad, 1997; Havre and Skogestad, 1998; Skogestad and Larsson, 1998; Skogestad et al., 1998) proposed selecting a set of controlled variables based on the concept of self-optimizing control and applied the concept to a number of systems. Downs (1992) emphasized the importance of inventory control. Ng and Stephanopoulos (1998) proposed a hierarchy for synthesizing a control structure by successively refining the flowsheet structure. A number of researchers (McAvoy and Ye, 1994; Lyman and Georgakis, 1995; Kanadibhotla and Riggs, 1995; Ricker and Lee, 1995; McAvoy et al., 1996; Ricker, 1996) have designed plantwide control systems for the Tennessee-Eastman process (Downs and Vogel, 1993).

Due to the large number of variables involved for the plantwide control synthesis problem, and a combinatorial growth in the total number of possible control structures with respect to the number of variables, a complete dynamic evaluation of all alternative control structures is impractical for

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any realistic process. For example, Price and Georgakis (1993) have shown that more than 70 possible control structures are possible for a simple reactor-separator-recycle system using only P/PI controllers and assuming a fixed set of controlled variables. Rather than evaluating all the possible alternatives, we would like to decompose the problem into a hierarchy of decisions. The decisions that have the most economic impact are made earlier in the hierarchy. At each level of hierarchy, alternatives are generated and only economically attractive alternatives are kept for further considerations. This procedure is motivated by Douglas's hierarchical procedure for conceptual process design (Douglas, 1988). In contrast to some other plantwide control procedures, decisions are made based on economics. By focusing on these economic trade-offs first, we also eliminate numerous economically unattractive control structures.

A brief description is presented of the basic idea behind the proposed procedure, and the reactor-separator-recycle system is described. The proposed procedure is then applied to this simple system. Future research work is outlined.

Basic Ideal Behind the Hierarchical Procedure

In general, the objective is to synthesize a plantwide control system that optimizes some measure of profitability. The profit (P) is defined as the difference between the revenues (R) and the total operating cost (C_{total}) (Douglas, 1988), that is

$$P \triangleq R - C_{\text{total}} \tag{1}$$

For each process alternative, the total operating cost is the sum of the raw materials and utility costs (C_{RU}), the annualized capital cost (C_{Cap}), the labor costs (C_{Labor}), and the annualized cost for the control system hardware/software (C_{CS}). Since a control system for an existing plant is being synthesized, the annualized capital cost is assumed to be constant. Thus, the following holds.

$$\max_{\text{plantwide control system}} P \Leftrightarrow \max_{\text{plantwide control system}} [R - C_{\text{RU}} - C_{\text{Labor}} - C_{\text{CS}}]$$

Clearly, this optimization problem must be solved subject to many constraints (such as production rate, equipment, safety, and so on). Also, the profit depends on the disturbances d. Thus

$$(\mathbf{PO}) \quad \max_{\text{plantwide control system}} [R - C_{\text{RU}} - C_{\text{Labor}} - C_{\text{CS}}](d) \quad \text{subject to constraints}$$

This optimization problem is generally too complex to be solved directly because of the need for an accurate dynamic model and the large number of decision variables involved (such as control/controller structures and tuning parameters). Decomposing the problem into a hierarchy of decisions is proposed, which is summarized in Table 1 and briefly discussed below. Proceeding down the hierarchy, more *modeling*, not structural, details are added to a control structure. A systems viewpoint is taken (that is, the complete flowsheet is considered at each decision level). The approach here differs from that of Ng and Stephanopoulos (1998) in that the structure of the flowsheet remains the same throughout our hierarchy but is successively refined throughout their hierarchy.

Table 1. Hierarchical Procedure for Plantwide Control

Step Descriptions	Mathematical Problem
1. Steady-State Robust Feasibility	Are the constraints feasible?
2. Robust Optimality—Controlled Variable Selection	$\max_{\text{controlled variables}} (R - C_{\text{RU}})^{SS}$
3. Steady-State Control Structure Screening	*RGA*, *SVD*, etc.
4. Dynamic Control Structure Synthesis	$\max_{\text{dynamic control structure}} (R - C_{\text{RU}})^{\text{Dyn}}$
5. Economic Ranking	$R - C_{\text{RU}} - C_{\text{Labor}} - C_{\text{CS}}$
6. Dynamic Simulations	

Step 1: Steady-State Robust Feasibility (Flexibility). Before a plantwide control system is synthesized, it is necessary to determine that the design is feasible for *all* the expected disturbances at steady state. Essentially, this step ensures that the constraints for the optimization problem (**P0**) are feasible at steady state.

Step 2: Controlled Variables Selection. Let

$$(R - C_{\text{RU}}) \triangleq (R - C_{\text{RU}})^{SS} + (R - C_{\text{RU}})^{\text{Dyn}}$$

where $(R - C_{\text{RU}})^{SS}$ denotes the "profit" if the plant were operated at steady state, and $(R - C_{\text{RU}})^{\text{Dyn}}$ accounts for the "profit" due to dynamic variations. Since the first term generally dominates and since evaluating the second term requires a dynamic model, the second term is ignored at this stage. If it is assumed that C_{Labor}, and C_{CS} are essentially the same for all control system alternatives, then the optimization problem (**P0**) becomes the following

$$\max_{\text{plantwide control system}} (R - C_{\text{RU}})^{SS}(d) \text{ subject to constraints}$$

which is equivalent to

$$(\mathbf{P1}) \quad \max_{\text{control variables}} (R - C_{\text{RU}})^{SS}(d) \quad \text{subject to constraints at steady state}$$

This optimization problem is much simpler to solve than the optimization problem **P0** as it is only necessary to determine a set of controlled variables *and* their set points; control/controller structures, controller tuning parameters, and process dynamics do not have to be considered. The set points may depend on *measured* or estimated disturbances, but not on unmeasured disturbances that are not estimated. The key idea in solving this optimization problem, discussed below, is to select a set of controlled variables and their set points so that the steady-state operation is close to being optimal for all the unmeasured disturbances. Solving this optimization

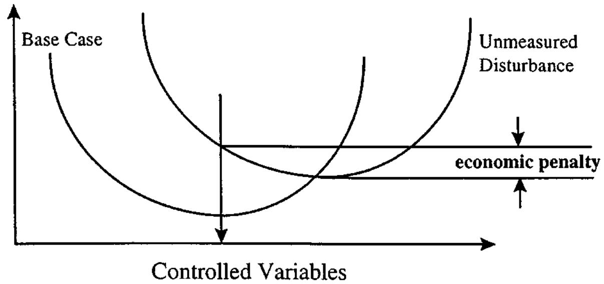

Figure 1. Economic penalty.

problem also quantifies the economic incentive for measuring or estimating a disturbance.

Given n steady-state degrees of freedom for a flowsheet, it usually does not matter which n independent variables are chosen to be fixed in the base case, as the solution is usually (but not always) unique. (The number of possible choices is usually enormous for any realistic problem.) However, disturbances always exist during plant operation. Optimal operation at the base case does *not* imply optimal operation when a disturbance enters. It is preferable to select a set of controlled variables which is insensitive, as far as economics are concerned, to all the unmeasured disturbances (that is, robust optimality). Mathematically, the following optimization problem is solved

$$\textbf{(P1}'\textbf{)} \quad \max_{\text{controlled variables}} \left\{ (R - C_{\mathrm{RU}})^{SS}(d, r) - (R - C_{\mathrm{RU}})^{SS}[d, r^*(d)] \right\} \text{ subject to SS constraints}$$

where r represents the optimal set points for a set of controlled variables when $d = 0$ and r^* = the optimal set points for the controlled variables if d were measured (and thus r^* depends on d). Alternatives can be generated and ranked according to the *economic penalties* (that is, the differences between optimal and actual steady-state costs; see Figure 1). Poor choices would be eliminated from further consideration. It should be noted that d includes the measurement errors for the controlled variables. Since an exact model is rarely available, we may want to solve **P1′** with respect to model uncertainty in addition to the disturbances d.

The idea of selecting controlled variables with a small economic penalty is not new. Findeisen et al. (1980) first introduced the concept of "feedback optimizing control." Arkun (1978) used this concept for the selection of controlled variables and applied the concept to a gasoline polymerization plant. Model uncertainty and measurement noise were not considered, however. Skogestad and co-workers extended the work to include the implementation error associated with the controlled variables and used the term "self-optimizing control" to describe the concept. The concept of "partial control" suggested by Arbel et al. (1996) can be seen as a variation of "feedback optimizing control." It should be noted that, while it is conceptually straightforward to solve **P1′**, it is difficult practically because of its combinatorial nature, especially in the presence of model uncertainty.

Step 3: Steady-State Control Structure Screening. Manipulated variables are selected for each set of controlled variables and alternative control structures are generated. Again, the total number of alternatives thus generated is usually very large. Qualitative controllability measures such as relative gain array (RGA), singular value decomposition (SVD), and so on, may be used to eliminate *poor* alternatives from further consideration. The rationale here is that the control structures with undesirable steady-state controllability measures may have small values of $(R - C_{\mathrm{RU}})^{\mathrm{Dyn}}$ and can be eliminated. This step also allows the designer to eliminate poor control structures based on experience.

Step 4: Dynamic Control Structure Synthesis. So far, only steady-state information has been used and the inventory and levels have been assumed to be perfectly controlled. Now, using the available (or additional) surge capacities to improve dynamic performance by smoothing the dynamic behavior of the process may be desired. Feedforward and cascade control structures can also be synthesized. For each set of controlled variables/manipulated variables, a control system is synthesized so that the following is optimized

$$\textbf{(P2)} \quad \max_{\text{dynamic control structure}} (R - C_{\mathrm{RU}})^{\mathrm{Dyn}}(d) \qquad \text{subject to constraints}$$

The Controllability Index ν introduced by Zheng and Mahajanam (1999) can be used to compute $(R - C_{\mathrm{RU}})^{\mathrm{Dyn}}$. The basic idea behind this index is as follows: What is the minimum *additional* surge capacity required for a given flowsheet and a given control system so that all the objectives and constraints are met *dynamically* for all the expected disturbances? Since the surge capacities introduced do not affect the steady-state behavior, they have no impact on the decision made in the previous steps.

Step 5: Economic Ranking. At this point, a set of alternative plantwide control structures has been generated. Based on the profit for each plantwide control structure, they are ranked according to economics.

Step 6: Dynamic Simulations. Finally, candidate plantwide control structures are verified via dynamic simulations.

For the remainder of the article, the hierarchical procedure described above is illustrated on a simple reactor-separator-recycle system. Some key issues are discussed, and practical solutions are proposed.

Reactor-Separator-Recycle System

Figure 2 depicts the simple reactor-separator-recycle (RSR) system studied by Papadourakis et al. (1987) and many other researchers. A first-order reversible, exothermic, and liquid-phase reaction $A \leftrightarrow B$ occurs in an ideal continuous stirred-tank reactor (CSTR). The reactor molar volume is 9,600 kmol, and the nominal values for the fresh feed stream are $F_0 = 0.2$ kmol/s, $T_0 = 300$ K, and $x_0 = 1$ (that is, pure A). For simplicity, it is assumed that the reactor is cooled by adjusting the cooling water exit temperature. The reactor rate expression is as follows

$$r_A = k_f e^{-E_f/RT} x_A - k_r e^{-E_r/RT}(1 - x_A)$$

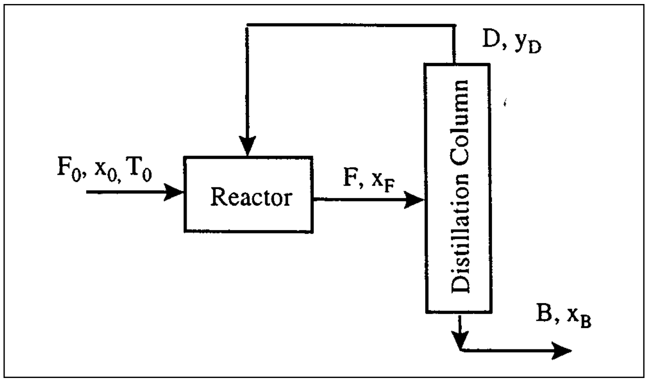

Figure 2. RSR system.

where x_A is the mole fraction of A and the kinetic parameters are given in Table 2. The reactor effluent is fed to a distillation column, where the lighter component A, which comes off as the distillate, is condensed and recycled to the reactor. Product B is obtained at the bottom. The column has 41 stages (including a partial reboiler) and a total condenser and the feed stage is the 20th stage from the top. For simplicity, theoretical trays, constant relative volatility of 1.5, constant molar overflow in the column, and saturated liquid feed are assumed. The holdups for all the trays, reboiler, and reflux drum are 12 kmol. Rigorous dynamic models of the reactor and the column (given in Skogestad and Postlethwaite (1996)) are used to simulate the different control structures and to evaluate their dynamic performance.

Here, three disturbances are considered: one measured and two unmeasured. The measured disturbance is that the fresh feed rate F_0 is expected to vary $\pm 30\%$ of its nominal value (corresponding to changes in the production rate). The two unmeasured disturbances are $\pm 30°K$ in fresh feed inlet temperature (T_0) and a -10% change in fresh feed composition (that is, change in $[x_{0A}, x_{0B}]$ from [1,0] to [0.9, 0.1]). The product purity specification is $99 \pm 0.2\%$.

This system has been studied by many researchers. For example, Papadourakis et al. (1987) concluded that the (steady-state) RGA for single units differs significantly from that in a plantwide environment. It was found that a control configuration that worked well for the column alone did not necessarily work well for the whole system. Price and Georgakis (1993) did exhaustive dynamic simulations on this system to develop guidelines for plantwide control system design and throughput manipulation in plantwide control structures. It was emphasized that a "self-consistent" control structure should be developed and those variables which preserved the plantwide nature of the problem should be selected for control. Luyben and co-workers studied the "snowball" effect (Luyben, 1994), "inventory control in processes with recycle" (Belanger and Luyben, 1997), "capacity-based economic approach" (Luyben and Elliot, 1995), and other concepts on this simple system. Wu and Yu (1996) have developed two different "balanced control schemes" for this system, to overcome the snowball effect and to achieve better disturbance rejection. They stressed the importance of proper "load distribution" among units. Luyben and Floudas (1992) illustrated the interaction of design and control for this system. This was done by incorporating steady-state economics and open-loop controllability measures within a multiobjective mixed-integer nonlinear program (MINLP) problem. They found that leaving the reactor and recycle stream compositions as variables in the NLP significantly reduces the cost when compared with designs where these compositions are fixed.

Table 2. Kinetic Parameters and Heat-Transfer Coefficient (hA)

Parameter	Value
k_f, s^{-1}	108
E_f/R, K	4,400
k_r, s^{-1}	10,050
E_r/R, K	5,912
$-\Delta H_{r\times n}$, kJ/kmol	12.5
hA, kJ/K·s	16,700
C_p, kJ/kmol·K	8

Application of the Hierarchical Procedure to the RSR System

Step 1: steady-state robust feasibility

This step checks if the system is feasible at steady state for all the expected disturbances. A steady-state model, as well as the sizes of all the expected disturbances, are needed to carry out this step. Suppose that the steady-state behavior of a process is described by the following n_f algebraic equations

$$f(u, y, d) = 0 \tag{2}$$

where $u \in \mathfrak{U} \subset \Re^{n_u}$ is the input (that is, manipulated variables) to the process, $y \in \mathfrak{Y} \subset \Re^{n_y}$ is the output, and $d \in \mathfrak{D} \subset \Re^{n_d}$ is the disturbance. $\mathfrak{U}$, $\mathfrak{Y}$, and $\mathfrak{D}$ are the sets of allowable inputs, allowable outputs, and expected disturbances, respectively. The steady-state robust feasibility problem is posed mathematically as follows:

Problem 1. Does there exist some $u \in \mathfrak{U}$ and $y \in \mathfrak{Y}$ such that Eq. 2 is feasible for each $d \in \mathfrak{D}$?

While this problem can be solved mathematically in a straightforward manner, a simpler approach is proposed by examining the feasibility of several common process design constraints.

Pressure. Pressure drops are necessary for flows in the right directions. It is important to install pumps, compressors, and so on, with necessary power to provide sufficient pressure drops in the worst-case situation.

Material Balance (Inventory Control). An "exit" point is needed for each component in the system. Such an exit point includes product stream, purge, or reaction. Particular attention should be paid to the trace components in the system (Joshi, 1990).

Energy Balance. Furnace size and utility streams must be sufficiently large so that energy can be balanced throughout the process in the worst case.

Equipment Capacity. Each individual unit must have sufficient capacity to handle the worst-case situation. This problem is coupled as some loading for one unit can be shifted to another unit.

RSR System. For the reactor-separator-recycle system, pumps with necessary power need to be installed at appropriate locations, depending on the operating pressures of the reactor and distillation column. If the feed contains inert impurities lighter than A, then a purge stream must be installed on the recycle stream. If impurities lighter than A are produced via reaction, then, depending on the kinetics, either a purge stream must be installed or sufficient reactor and column capacities need to be provided. A purge stream must be installed if the reaction is irreversible. As far as the energy balance is concerned, cooling and heating capacities for the reactor and column must be sufficient to handle the worst-case situation, in this case, a 30% increase in production rate and a 30°K decrease in feed inlet temperature. The reactor and the distillation column should have the capacities to handle the worst-case situation. For simplicity, it is assumed that both the reactor and the distillation column have been overdesigned so that their equipment constraints are not violated in the worst case.

Step 2: controlled variable selection

According to Eq. 2, the steady-state number of degrees of freedom equals $n_D = n_u + n_y - n_f$, where n_u is the number of inputs, n_y is the total number of outputs, and n_f is the total number of equations, which implies that n_D variables need to be fixed to have a unique steady-state solution. Notice that fixing n_D variables is necessary, but not sufficient, to guarantee a unique steady-state solution. The question is: Which n_D variables should be chosen? It is worthwhile to point out that the controlled variables can be a combination of inputs (that is, u) and outputs (that is, y). The set of controlled variables can be divided into economic (such as recycle flow) and noneconomic variables (such as product purity) (Fisher et al., 1988c). (While, theoretically, there is no need to distinguish between the economic and noneconomic variables, this distinction may be useful practically. Since it reduces the number of controlled variables that need to be chosen to minimize the economic penalty and since choosing an optimal set of controlled variables is a combinatorial problem, this distinction should simplify the computation.) The choice of the noneconomic variables is usually obvious from the problem statement. Several criteria need to be satisfied in selecting the controlled variables:

(1) They must ensure steady-state robust feasibility. Suppose that the set point for a set of controlled variables is r, which may depend on measured (or estimated) disturbances (d_m), but not on unmeasured disturbances that are not estimated (d_u), that is, $r(d_m)$. Such specifications provide n_D equations, that is

$$c(u, y, r(d_m), d) = 0 \tag{3}$$

Now the steady-state robust feasibility problem becomes the following:

Problem 2. Do there exist some $u \in \mathfrak{U}$ and $y \in \mathfrak{Y}$ such that the following set of equations

$$\begin{gathered} f(u, y, d) = 0 \\ c(u, y, r(d_m), d) = 0 \end{gathered} \tag{4}$$

is feasible for each $d \in \mathfrak{D}$?

Notice that the feasibility of *Problem 1* is necessary, but not sufficient, for the feasibility of Problem 2. The infeasibility can be caused by ill-chosen controlled variables (that is, the problem is infeasible regardless of their set points) or ill-defined set points (that is, the problem may be feasible for some properly chosen set points). The latter is illustrated by the following example.

Example 1. The overall material balance for the RSR system is considered. The amount of the product produced must equal the amount of the feed reacted (that is, $B = F_0 = Vr_A$). Substituting the expression for r_A and $V = 9{,}600$ kmol

$$F_0 - 1.04\times 10^6 e^{-4{,}400/T} x_A + 9.65\times 10^7 e^{-5{,}912/T}(1 - x_A) = 0$$

where $0.14 \le F_0 \le 0.26$ and $0 \le x_A \le 1$. There exist values of T and x_A such that this equation is feasible for all allowable values of F_0 (that is, Problem 1 is feasible). However, depending on the choice of the controlled variable and its set point, Problem 2 may not be feasible. For example, if it is chosen to control x_A at 0.55, then the equation is not feasible for some allowable values of F_0 (such as 0.26), although it is feasible for the base case $F_0 = 0.2$. However, if it is chosen to control x_A at 0.6, then the equation is feasible for all allowable values of F_0. Therefore, the infeasibility of Problem 2 is caused by an ill-defined set point.

(2) They should ensure robust optimality. In the nominal case (that is, $d = 0$), the steady-state economics do not depend on the set of controlled variables chosen, as long as it forms a basis and its set point corresponds to the optimal operating conditions. However, this is not the case in the presence of an unmeasured disturbance. When an unmeasured disturbance enters the process, the process is no longer operating at its optimal operating conditions. The difference in operating costs is called the economic penalty; Figure 1 illustrates the basic idea. The objective in selecting the set of controlled variables is to minimize the economic penalty, either in the worst case or on the average. (Based on our limited experience, there is little difference between the two objectives. The objective should be chosen based on the simplicity of the resulting optimization problem.) Furthermore, the economic penalty should be insensitive to the actual values of the set point (that is, a "flat valley" is preferred). This is because of dynamics and potential measurement errors.

With these criteria in mind, the alternatives are generated according to the following steps:

(1) Determine the number of steady-state degrees of freedom (that is, n_D) and the optimal operating conditions for each set of measured disturbances.

(2) Determine the noneconomic variables, denoted by y_{ne}, and their set points. Check for feasibility of Problem 2 (that is, Eq. 4).

(3) Generate alternative sets of economic controlled variables ($n_D - n_{y_{ne}}$ of them).

(4) Eliminate those alternatives which are not feasible (that is, Problem 2 has no solution).

(5) Since perfect control may not be possible (such as measurement errors and dynamics), a sensitivity analysis should be carried out to determine how sensitive the total annual-

Table 3. Optimal Operating Conditions at Various Production Rates*

Variable	Base Case	+30% in F_0	−30% in F_0
Fresh feed flow, F_0 (kmol/s)	0.20	0.26	0.14
Recycle flow, D (kmol/s)	0.31	0.43	0.2
Reactor effluent flow, F (kmol/s)	0.50	0.69	0.33
Column bottoms flow, B (kmol/s)	0.20	0.26	0.14
Distillate comp., x_D (mol fraction)	0.899	0.905	0.886
Cooling water temp., T_c (K)	325	338	315
Reactor temp., T (K)	326.4	331	321
Reflux flow, L (kmol/s)	0.95	1.29	0.63
Vapor boilup, V (kmol/s)	1.26	1.71	0.83
V/B	6.31	6.63	5.94
L/D	3.11	3.01	3.21
Column feed comp., x_F (mol fraction)	0.548	0.568	0.522
TAOC (in $\times 10^6$ \$)	7.3	10.0	4.8

*TAOC equals the sum of the energy costs for the reactor and the column. An energy price of \$3.80/MM kJ, the heat of vaporization of 26,500 kJ/kmol for both components, and an annual operating time of 8,000 h are assumed.

ized operating cost (TAOC) is to the actual set point for the controlled variables.

(6) Rank the remaining alternatives according to economic penalties (either in the worst case or on the average).

Remark 1. For m disturbances, evaluating the worst-case economic penalty requires solutions of 2^m problems, assuming that the worst case occurs on a corner. This may not be practical for large values of m. One can use the following short-cut method to simplify computations as many of the disturbances may be insignificant economically (Arkun, 1978; Fisher et al., 1988c): Classify a disturbance either as insignificant or dominant, depending on the economic penalty incurred when only this disturbance affects the system. Only the dominant disturbances will be used to determine the worst-case economic penalty. In most situations, this analysis should be adequate.

The RSR System. Now the above procedure is applied to the reactor-recycle-separator system.

(1) The system at steady state has 3 degrees of freedom, assuming that properties for the feed stream are fixed and that the column pressure is fixed.

(2) There is only one noneconomic variable and that is the product purity (99%). Problem 2 would not be feasible if the feed contained more than 1% of components heavier than B or more than 1% were produced via reaction.

(3) We need to select two more controlled variables. Suppose that there are six candidate variables to choose from and they are y_D, T_c, T, RR, L and D. (These six variables were chosen based on engineering intuition and common control practice. Clearly, more can be added, for example, the feed to the column.) Notice that the bottoms flow is not a candidate since it equals F_0 at steady state. The total number of alternatives is 15. The optimal operating conditions are determined for the three production rates (Table 3). The set points depend on the production rate since the production rate is a measured disturbance.

(4) The economic penalties for the fifteen alternatives are shown in Table 4. Notice that four of them (such as the $\{RR, L\}$ configuration) are not feasible even for the base case and that some of them (such as $\{T_c, L\}$) are not robustly feasible.

(5) When the reflux ratio (RR) and reactor temperature (T) are chosen as controlled variables (in addition to the product composition), the sensitivities of the TAOC to T and RR are shown in Figure 3. This sensitivity analysis allows one to determine how tightly each controlled variable should be controlled and how accurately each controlled variable should be measured, as far as the economics is concerned. If this requirement is too strict, we should eliminate the set of controlled variables from further consideration, at least in the initial stage of the design.

(6) Based on the results in Table 4, the following four alternatives are explored further. (Both the worst case and average case yield the same results.) In all the cases, the bottoms composition is controlled.

Case 1. T, RR
Case 2. T, y_D
Case 3. T, L
Case 4. T, D

Step 3: steady-state control structure screening

In this step, alternative control structures are synthesized for each set of controlled variables. Poor control structures

Table 4. Economic Penalties (%) for Various Combinations of Disturbances

CVs	$F_0 = 0.20$		$F_0 = 0.26$		$F_0 = 0.14$		Avg.
	+10% in T_0	−10% in T_0	+10% in T_0	−10% in T_0	+10% in T_0	−10% in T_0	
RR, T_c	4.04	7.87	4.28	13.96	5.00	3.74	6.48
RR, T	0.76	0.68	0.52	0.76	1.09	0.65	0.75
T, y_D	0.76	0.69	0.53	0.81	1.13	0.68	0.77
T_c, y_D	4.00	9.52	4.18	Infeasible	3.97	3.36	Infeasible
D, T	0.78	0.69	0.55	0.76	1.10	0.66	0.76
T, L	0.76	0.69	0.52	0.76	1.31	0.66	0.78
D, T_c	4.20	7.22	4.62	12.29	3.95	2.94	5.87
T_c, L	4.15	Infeasible	4.22	Infeasible	4.22	4.44	Infeasible
D, y_D	0.76	Infeasible	0.52	Infeasible	1.10	Infeasible	Infeasible
RR, y_D	0.76	Infeasible	Infeasible	Infeasible	Infeasible	Infeasible	Infeasible
L, y_D	0.76	Infeasible	Infeasible	Infeasible	Infeasible	Infeasible	Infeasible
RR, L	Infeasible	Infeasible	Infeasible	Infeasible	Infeasible	Infeasible	Infeasible
RR, D	Infeasible	Infeasible	Infeasible	Infeasible	Infeasible	Infeasible	Infeasible
T, T_c	Infeasible	Infeasible	Infeasible	Infeasible	Infeasible	Infeasible	Infeasible
L, D	Infeasible	Infeasible	Infeasible	Infeasible	Infeasible	Infeasible	Infeasible

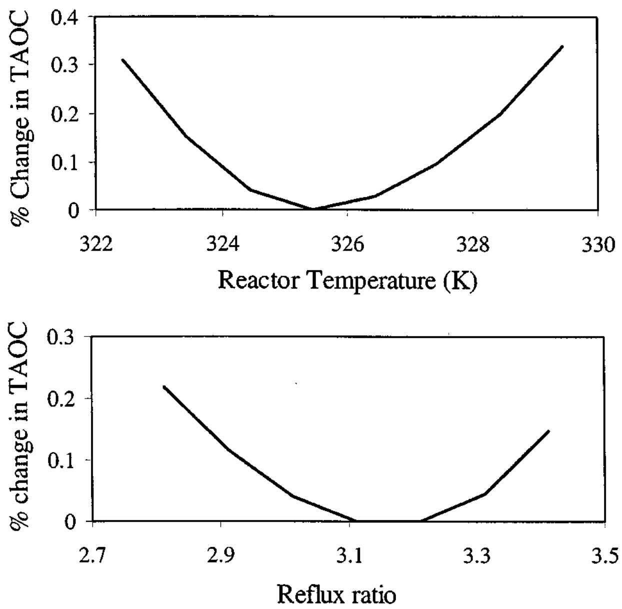

Figure 3. Sensitivity of TAOC to reactor temperature (T) and reflux ratio (RR): Case 1.

can be eliminated based on qualitative steady-state controllability indexes such as RGA and SVD. The rationale behind this step is to significantly reduce the potentially large number of alternatives based on the assumption that the control structures with undesirable steady-state controllability measures will likely have small $(R-C_{\mathrm{RU}})^{\mathrm{Dyn}}$. For each set of controlled variables, it is necessary to select manipulated variables and, if a decentralized control structure is used, pair the manipulated variables and the controlled variables.

RSR System. Manipulated variables are selected for each set of controlled variables from the available inputs. RGA is used to eliminate poor alternatives from further consideration. For the four most economically attractive sets of controlled variables determined in Step 2, the manipulated variables and RGA results are summarized below:

Case 1 (T, RR, x_B). Since RR is flow controlled, two possible manipulated variables are reactor cooling water temperature T_c and vapor boilup (V). The RGA value of 0.996 indicates that T should be controlled by T_c and x_B by V.

Case 2 (T, y_D, x_B). One possible set of manipulated variables is (T_c, L and V). The RGA is found to be dependent on steady-state operating conditions. The RGA matrices, at a feed rate (F_0) of 0.20 kmol/s, 0.26 kmol/s, and 0.14 kmol/s, are

$$\begin{bmatrix} 1.003 & 0.002 & -0.005 \\ -0.002 & 0.839 & 0.162 \\ -0.002 & 0.159 & 0.843 \end{bmatrix}, \quad \begin{bmatrix} 1.001 & 0.002 & -0.003 \\ -0.002 & 0.980 & 0.021 \\ 0.001 & 0.018 & 0.981 \end{bmatrix},$$

$$\begin{bmatrix} 0.775 & -0.103 & 0.328 \\ 0.001 & 1.012 & -0.013 \\ 0.224 & 0.091 & 0.685 \end{bmatrix}$$

respectively. (The rows and columns of the RGA matrices may not add up to one due to rounding.) The above RGA matrices suggest the following pairing: $T_c - T$, $L - y_D$, $V - x_B$. Notice that the above RGA matrices are determined based on the whole flowsheet.

Case 3 (T, D, x_B). With D being flow controlled, the manipulated variables are T_c and V. The RGA value of 0.996 indicates that T should be controlled by T_c and x_B by V.

Case 4 (T, L, x_B). Since L is flow controlled, the manipulated variables are T_c and V. The RGA value of 1.005 indicates that T should be controlled by T_c and x_B by V.

Step 4: dynamic control structure synthesis

The control alternatives synthesized so far have used only the steady-state information and are incomplete. It is necessary to synthesize the following three control structures, which affect $(R-C_{\mathrm{RU}})^{\mathrm{Dyn}}$ but not $(R-C_{\mathrm{RU}})^{SS}$:

(1) Level Control Structure
(2) Cascade Control Structure
(3) Feedforward Control Structure.

In this section, how the level control structure is synthesized is only discussed. Also, it should be noted that $(R-C_{\mathrm{RU}})^{\mathrm{Dyn}}$ depends on controller dynamics.

Our philosophy for designing level control systems is to use the available surge capacity to "smooth-out" dynamic fluctuations in flows, compositions, and so on. The design guidelines use Morari's proximity rule (Morari, 1983) and the impacts of flow disturbances on performance. Morari's proximity rule suggests to pair a manipulated variable which is physically "close" to the controlled variable; physical closeness usually implies short time constants and small dead times and, thus, superior control performance. Those flows that have the most significant impacts on performance should be chosen to control the levels, or more precisely, the available surge capacities should be used to minimize the dynamic fluctuations for those flows. Thus, the emphasis is *not* on tight control of levels.

The cascade control structure is used to minimize the effects of disturbances and model uncertainty, while the feedforward control structure is used to minimize the effect of measured disturbances. While qualitative guidelines exist in the literature on how to synthesize these systems, quantitative guidelines, based on economics, need to be developed.

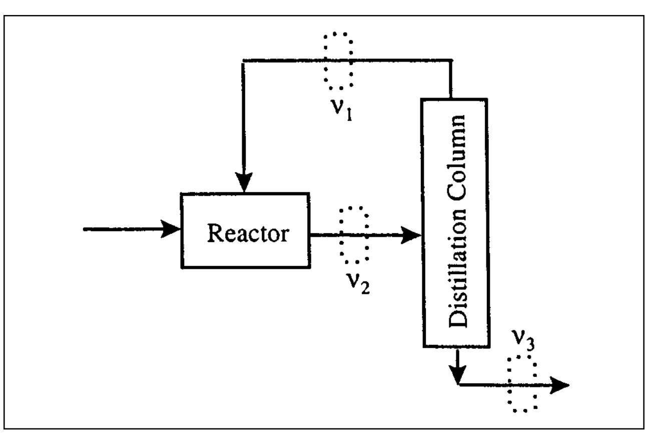

Figure 4. Determination of ν for the RSR system.

Table 5. Level Control Strategies for Case 1 Assuming a Fixed Production Rate Control

No.	Production Rate	Reactor Level	Reflux Drum Level	Reboiler Level
1	F_0	D	F	B
2	F_0	D	B	F
3	F_0	F	D	B
4	F_0	F	B	D
5	F_0	B	F	D
6	F_0	D	B	F

The RSR System. There are three levels to be controlled: the reflux drum, the reboiler, and the reactor levels. Notice that the reactor level differs from the other two levels as it affects the steady-state behavior and that it may have to be controlled more tightly. There is an additional level associated with the production rate: It is imagined that the product flows into a storage tank, which is taken out according to the demand (whose average value is called the production rate). Maintaining the precise value of production rate dynamically is not the objective; rather, the objective is to maintain the level in the product storage tank. Thus, there are four levels that need to be designed for each steady-state control alternative. Many alternatives are possible. For example, Case 1 is considered. For simplicity of discussion, F_0 is used to control the product storage tank level (production rate). Since the vapor boilup has already been chosen for the control of product composition, there are four flows (reactor effluent, reflux, distillate, and bottoms), out of which three *independent* flows need to be chosen to control the remaining three levels. Since the reflux ratio is controlled, the reflux and the distillation flows cannot be used simultaneously for controlling the levels. Thus, there are a total of six alternatives (Table 5). Using Morari's proximity rule, the following level control strategy is obtained: D for the reflux drum level, F for the reactor level, and B for the reboiler level.

By following the similar arguments for each set of controlled variables, the following four plantwide control structures are obtained:

$$CS\#1 \quad T_c - T,\ RR - RR,\ V - x_B,\ D - \text{reflux drum level},$$
$$B - \text{reboiler level},\ F - \text{reactor level}.$$
$$CS\ \#2 \quad T_c - T,\ L - y_D,\ V - x_B,\ D - \text{reflux drum level},$$
$$B - \text{reboiler level},\ F - \text{reactor level}.$$
$$CS\ \#3 \quad T_c - T,\ L - L,\ V - x_B,\ D - \text{reflux drum level},$$
$$B - \text{reboiler level},\ F - \text{reactor level}.$$
$$CS\ \#4 \quad T_c - T,\ D - D,\ V - x_B,\ L - \text{reflux drum level},$$
$$B\text{-reboiler level},\ F - \text{reactor level}.$$

Their PI&D's are shown in Figure 5, and the controller tuning parameters are given in Table 6. Notice that the feedforward control strategy of adjusting the set points according to the production rate is not shown in Figure 5 for simplicity. In all the cases, the production rate is controlled by F_0.

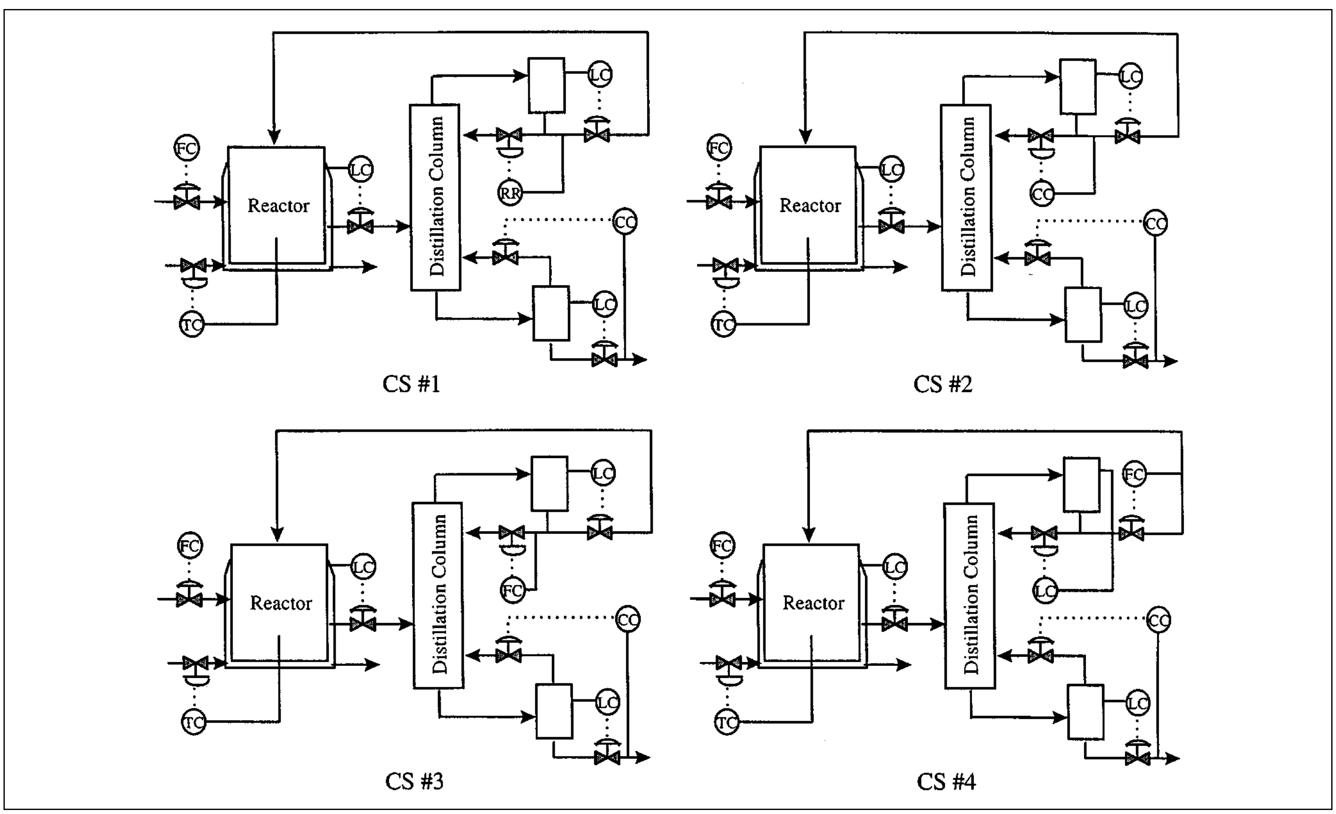

Figure 5. PI&Ds for the four control structures.

Table 6. Controller Tuning Parameters*

Controller	CS No. 1	CS No. 2	CS No. 3	CS No. 4
$T_c - T$	$k=1$, $\tau=20$	$k=1$, $\tau=20$	$k=1$, $\tau=20$	$k=1$, $\tau=20$
$V - x_B$	$k=-3.2$, $\tau=16.3$	$k=-0.41$, $\tau=23.4$	$k=-0.41$, $\tau=23.4$	$k=-0.41$ $\tau=23.4$
$L - y_D$	NA	$k=0.0412$ $\tau=18.9$	NA	NA
Reflux drum level	$k=1$	$k=10$	$k=10$	$k=10$
Reboiler level	$k=1$	$k=10$	$k=10$	$k=10$

*The form of PI controller is $k(1+1/\tau s)$, τ in min, and the reactor level and the column pressure (and RR if applicable) are assumed to be perfectly controlled.

Now that alternative plantwide control systems have been generated, the question is: Which one of them is the most attractive *economically*? Zheng and Mahajanam (1999) recently introduced a simple controllability index ν to quantify the cost associated with the dynamic controllability [that is, $-(R-C_{\mathrm{RU}})^{\mathrm{Dyn}}$]. ν is defined to be the minimum (additional) surge capacity necessary to meet all the objectives dynamically for all the expected disturbances. For a continuous process whose product quality specifications are one-dimensional (for example, purity), ν is finite if the steady-state problem is feasible and the closed-loop system is asymptotically stable. ν depends on the process dynamics, process constraints, product variability, disturbance characteristics, and controller/control structures. It is based on the belief that poor quality control can be overcome by installing sufficiently large surge capacities in the process. If no additional surge capacity is needed, then there is no cost associated with dynamic controllability.

For a continuous process with N_s streams, a surge tank of volume ν_i is installed for stream i. This is illustrated for the RSR system (Figure 4). Then, the controllability index equals the sum of the optimal values of ν_i, that is

$$\nu = \sum_{i=1}^{N} \nu_i^{\mathrm{opt}}$$

where ν_i^{opt} is the solution to the following optimization problem

$$\min_{\nu_i} \ \Phi(\nu_i)$$

$$\text{subject to} \begin{cases} y(t) \in \mathfrak{Y} \ \forall t \\ d(t) \in \mathfrak{D} \ \forall t \\ u(t) \in \mathfrak{U} \ \forall t \\ \text{process dynamics} \\ \text{controller dynamics} \end{cases} \tag{5}$$

The objective function $\Phi(\nu_i)$ depends on the particular application. For example, $\Phi(\nu_i)=\sum_{i=1}^{N}\nu_i$ minimizes the total volume while $\Phi(\nu_i)=\sum_{i=1}^{N} f_i(\nu_i)$, where $f_i(\nu_i)$ is the annualized cost associated with installing and maintaining the surge tank of volume ν_i, minimizes the total cost. The process dynamics can be represented by a process model and its associated uncertainty. Also, for existing processes where surge volumes may not be removed, it is necessary to impose additional constraints $\nu_i \geq 0$.

RSR System. Clearly, to compute ν for a plantwide control system, the allowable product variability and disturbance characteristics must be specified. The cost associated with dynamic controllability for the four plantwide control structures are given in Table 7. Here ν_1, ν_2, ν_3 represent the size of surge tanks installed on the recycle, reactor effluent, and product streams (Figure 4). The product composition specifications are $99 \pm 0.2\%$ and disturbances are assumed to be step-like for simplicity. A cost correlation $5{,}630\nu^{0.65}$, obtained from Peters and Timmerhaus (1980), is used for computing the annualized cost of installing and maintaining a surge tank of volume ν. Notice that this correlation does not consider the inventory and environmental costs of the material inside the tank, which can be significant in many cases. It should be emphasized that the surge capacities depend on the controller dynamics (tuning parameters). It has not been attempted to optimize the surge capacities over controller tuning parameters. A reasonable set of controller tuning parameters have been merely chosen. However, this optimization, if necessary, can be done.

Step 5: economic ranking

Up to this point, alternative plantwide control systems have been synthesized and for each of them $(R-C_{\mathrm{RU}})^{SS}$ and $(R-C_{\mathrm{RU}})^{\mathrm{Dyn}}$ have been computed. Once C_{Labor} and C_{CS} for each alternative have been estimated, the annual profit for each plantwide control system is computed and ranked economically. If C_{Labor} and C_{CS} are assumed essentially the same for all the control structures, then CS No. 1 corresponds to the optimal plantwide control system. Notice that the optimal plantwide control system does not use a dual composition control strategy on the column. It is worthwhile to emphasize again that the optimal plantwide control system depends on the set of expected disturbances.

Step 6: dynamic simulation

Figure 6 shows dynamic simulation results of the four structures for the worst-case disturbance (30% increase in F_0; 10% increase in T_0). It is evident that CS No. 1 is better than

Table 7. Cost of Dynamic Controllability for Selected Plantwide Control Structures*

Control Structure	CS No. 1 Capacity (m^3)	CS No. 1 Cost ($)	CS No. 2 Capacity (m^3)	CS No. 2 Cost ($)	CS No. 3 Capacity (m^3)	CS No. 3 Cost ($)	CS No. 4 Capacity (m^3)	CS No. 4 Cost ($)
ν_1^{opt}	0	0	19	38,500	17	36,000	15	33,000
ν_2^{opt}	0	0	8.7	23,000	4.7	15,000	0	0
ν_3^{opt}	0	0	39	60,500	64	84,000	75	93,000
ν or total cost	0	0	66.7	122,000	85.7	135,000	90	126,000

*A molar density of 0.0232 m^3/kmol is assumed.

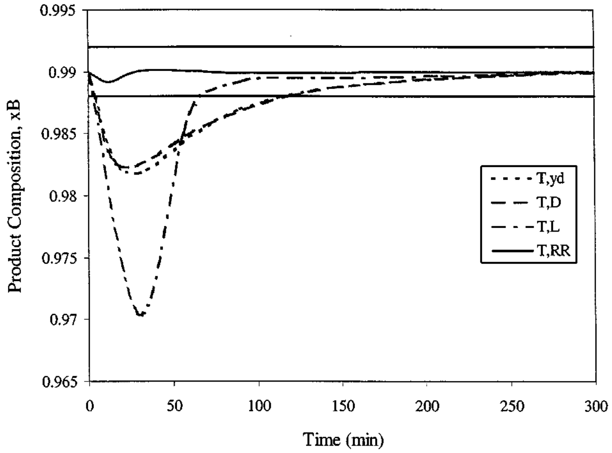

Figure 6. Dynamic simulation of four structures (CS No. 1: solid; CS No. 2: dotted; CS No. 3: dash-dot; CS No. 4: dashed) for the worst-case disturbance.

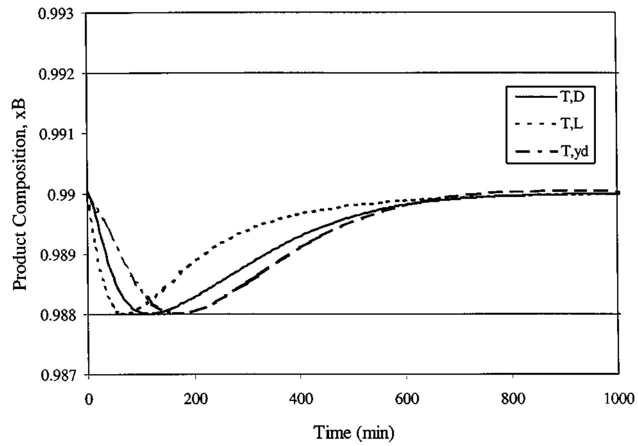

Figure 7. Responses of CS No. 4 (solid), CS No. 3 (dotted) and CS No. 2 (dash-dot) with additional surge capacities.

the other structures, since it satisfies the performance specifications at all times with no additional surge capacity. Figure 7 shows that CS Nos. 2, 3, and 4 satisfy product variability specifications after the additional surge capacities given in Table 7 have been installed.

Summary of Application of the Hierarchical Procedure to the RSR System

The decisions made at each step of the hierarchical procedure for the RSR system are shown in Figure 8. As one proceeds down the hierarchy, modeling details are added to the flowsheet and alternatives are generated. The idea is *not* to explore every alternative, but to quickly eliminate poor alternatives from further consideration. For an existing plant with an existing plantwide control system, one can move up the hierarchy and compare the economic incentive for redesigning the plantwide control system.

Conclusions

A hierarchical procedure for systematically synthesizing a plantwide control system has been proposed and its basic ideas have been illustrated on the simple reactor-separator-recycle system. The procedure was motivated by Douglas' conceptual design framework (Douglas, 1988) and the fact that many plantwide control alternatives exist for any realistic chemical process. The procedure decomposes the problem into a hierarchy of decisions. At each level of the hierarchy, decisions are made based on economics. The cost associated with dynamic controllability is quantified through the Controllability Index ν introduced by Zheng and Mahajanam (1999).

There are a number of practical issues that need to be resolved in carrying out the procedure. For example, the problem of selecting a set of controlled variables is a combinatorial problem and an efficient method needs to be developed; an approximate tool for estimating ν quickly needs to be developed; quantitative guidelines for deciding when feedfor-

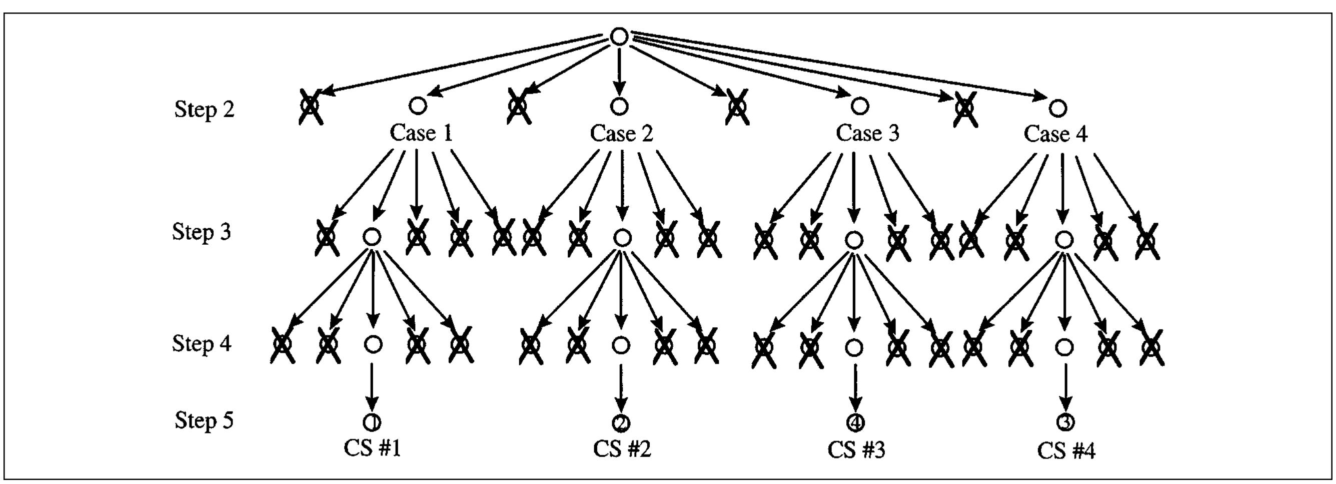

Figure 8. Decision tree for the RSR system.

ward and cascade structures are justified economically should be developed; and so on. Thus, it is expected that the procedure will provide a focal point for the definition of new research problems in plantwide control.

The procedure also lays the foundation for a framework for integrating process design and process control (such as designing a process with the minimum inventory). In this case, the annualized capital cost is no longer constant. At the design stage, developing a dynamic model is rarely justified due to both lack of data and developmental cost. Thus, it is important to have a procedure that can make decisions regarding controllability with only steady-state information. A valuable byproduct of the research should be an improved conceptual framework for the teaching of process control. Currently, most undergraduates are exposed to the design of a controller *given* a set of controlled and manipulated variables, so that they develop no intuition or experience with the decisions required to develop a plantwide control system.

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