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OPERABILITY OF INTEGRATED PLANTS

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Abstract

Various approaches exist for synthesizing plant-wide control, from heuristics to mixed-integer nonlinear programming (MINLP). Few specifically address the problems associated with heat and mass integration in processes. This paper extracts many integration specific aspects and discusses their effect on plant control. The focus is on limitations and tradeoffs within operability. Guidelines are presented for finding good control strategies for integrated plants. We hope this can initiate some work alleviating the engineers problems with this issue.

1 Introduction

Design of control systems is usually done by examining small portions of the overall process. This gives rise to multiloop control systems which have the advantage of simplicity in commissioning, startup, shutdown and tuning. Systems of this kind cover more than 90 % of the applied control in most plants today. It is of interest to assess the operability to ensure that proper operation is feasible for the integrated plant as a whole. With operability we mean that a plant is controllable, flexible, has an optimal operating point and that suitable actuators and measurements are chosen.

Increased energy efficiency has been a major force in recent research (Linnhoff and Hindmarsh, 1983), giving rise to plants which are more integrated. This integration of the process units may occur during initial design or at a later stage. The requirements for these two cases are different, since an add-on integration scheme does not usually have the same freedom to change process units to facilitate better operability. Thus, the results from these two approaches are interpreted based on different requirements. We will primarily consider operability of fixed designs.

The main problem in integrating design and control is the exponential growth in the number of alternatives as the problem size increases. First, steady state process design is itself a combinatorical problem. Second, to evaluate the controllability of a given design, one must consider the best of a number of alternative control structures, which again is a problem that rapidly increases with size. Thus, to integrate design and control one must first be able to, for a given process design, suggest a reasonable control structure and evaluate its control characteristics. In the literature review we have therefore also included references to work on plant wide control system design (structure selection).

Numerous authors have presented work that has tried to unify the design and control stage of process development. The wish to combine these tasks and the many facets of their implementation (heuristic, mathematical programming, analysis/judgement) have provided the multitude of approaches that exist today. As major directions they can be termed "concurrent", "sequential" or "iterative".

The objective of this paper is to illustrate the difficulties in analyzing the degrees of freedom for integrated plants. We also discuss the various tradeoffs and limitations in combining design and control for integrated plants. Finally, guidelines are presented to help identify operability deficiencies.

2 Previous work

This brief review covers the main trends within control of integrated plants and plant-wide control. The references are given chronologically to easier see the developments within the area.

Buckley (1964) proposed to divide plant-wide control into two levels; first material balance (MB) control followed by quality control. The justification relies on the fact that MB and quality control are on different time scales and thus independent of each other (MB - slow, Quality - fast). This separation is somewhat arbitrary and proves even less clear for integrated processes. Buckley has a somewhat pessimistic view on operability: "...it is usually necessary to go to a semiworks or pilot plant to demonstrate integrated operation, *process operability*, a final flowsheet and process economics."

Foss (1973) regards the fundamental problem in process control as not the development of more sophisticated control but rather a framework for selecting manipulated and measured variables for control. The same challenge exists today

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for plant wide control and control of integrated plants. Foss also claims that "it is the presence of coupling among many variables that is primarily responsible for the near inscrutable complexity of dynamic processes."

Rinard (1975) mentions operability in his paper and extends the discussion to *integrated* plants, where he anticipates more interaction, less independent manipulated variables and slower response than for un-integrated designs.

Umeda *et al.* (1978) present a logical structure for control system synthesis. The emphasis is on applying control to all process units individually followed by a coordination phase removing controllers until the system is not overspecified. Structural matrices and heuristics are used during the second phase.

Morari *et al.* (1980) present an approach to systematic control system design. Here the focus is on decomposition for optimal control.

Morari (1981) subscribes to the need for decomposition in control system synthesis and gives a good discussion on practical results. Structuring of the control tasks, selection of measured and manipulated variables, and interdependencies between design and control are viewed as demanding problems.

Nishida *et al.* (1981) state that "...the synthesis of control systems for complete chemical plants is a problem within a steady state environment," and claims that the most important development to come is identifying the interaction and finding the basic variables that will affect the structure of the plant. The dynamic aspects of both interaction and control have received a rising interest, while the identification problem still holds.

Govind and Powers (1982) propose synthesizing control systems using structural information (e.g. cause-effect graphs) and simple models.

Stephanopoulos (1982) gives a comprehensive review of problems in process control and claims that identification of manipulated variables, measurements and controller structure is "where novel and imaginative formulations are needed." Decomposition is emphasized and the article also treats issues such as fault detection and startup/shutdown.

Grossmann *et al.* (1983) defined operability as encompassing flexibility, controllability, reliability and safety although their focus is on using mathematical programming for generating flexible processes.

Calandranis and Stephanopoulos (1986) look at the interaction between operability and design of heat exchanger networks. Instead of generating a single criteria for operability they checked if the various networks were operable within a predetermined set of operating (including disturbance) parameters. This also allows for identifying what factors limit the realization of the desired operational range. The approach concentrates on detecting *in*operability at steady-state and is not very transferable to general process plants. The checking of corner points within the parameter space does not always suffice.

Georgakis (1986) proposes to use extensive variables to reduce interaction to one-way and simplify control design.

Fisher *et al.* (1988) evaluate both controllability and operability when looking at the interface between design and control. Their example contains process integration without this being an issue. The controllability analysis checks for enough manipulated variables and operability is considered as extent of overdesign.

Kravanja and Glavič (1989) have evaluated total heat integrated plants, but have only considered control to alleviate a specific problem; a cyclic feed disturbance.

Georgiou and Floudas (1989) use structural modeling to pose the control synthesis problem as a MILP. The framework for generating all feasible control structures is based upon the concepts of connectability, controllability and observability. Selection within structures is not addressed.

Perkins (1989) looks at the interaction between design and control by considering both fast and slow effects, which transforms to the assessment of controllability, switchability and flexibility, or summarized as operability. Linear indices are acknowledged (γ, $\underline{\sigma}$) as well as economic indicators.

Price and Georgakis (1992) give a good discussion on control of large scale systems. The article introduces the terms "self-consistency" and "primary process path" which gives additional understanding to Buckley's material balance control.

Morari (1992) reviews the effect of design on *controllability* and points out:

1. Steady state controllability criteria are unqualified for analyzing dynamic behavior.
2. Ad hoc controller tuning may quantify irrelevant design changes as attractive.
3. Better, simpler controllability criteria are needed before algorithmic synthesis techniques to trade of controllability and economics become meaningful.

Point 3 is valid in general while point 2 can be extended to design of integrated plants; the effect of a design change holds more significance when the flowsheet is reoptimized to account for this change.

Ponton and Liang (1992) adapt Douglas' (1988) stepwise design theory to control system design by designing the control system in parallel with the plant.

Schmidt *et al.* (1992) have investigated operability of energy integrated distillation columns. The heat-pump between reboiler and condenser reduced the operating space and gave instability. The instability was easily dealt with by control and did not have a major impact on controllability.

Bouwens and Kösters (1992) comment on the trend towards more integrated plants and recognize that operability is also important under non-steady state operating conditions.

Downs and Vogel (1993) present a challenge problem for studying and evaluating control technology. This example is well suited for studying plant wide and integrated control issues.

To summarize; there have been proposed many techniques for large scale process control, but there is still a great need for future work in this area.

3 Plant-wide control system design

3.1 "Top-down" selection of controlled and manipulated variables

The first step in designing a complete control system is often a degree of freedom (DOF) analysis. One generally starts from the top by identifying the overall control objectives and then moves towards the bottom of the control hierarchy. To check that there exist degrees of freedom (independent variables, valves, etc.), a "tick-off" procedure is useful, that is one identifies for each objective a degree of freedom.

There must be *at least* as many DOF (variables, valves, etc.) as there are control objectives to be met. For controllability reasons (speed of response) this requirement must also be satisfied locally. This gives another justification for per-

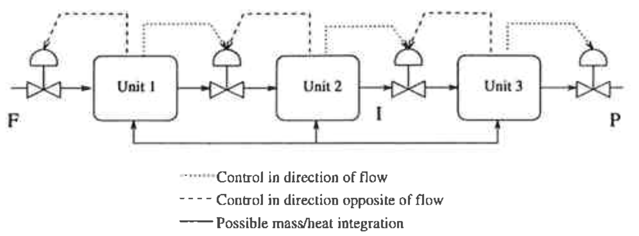


Figure 1: Production control schemes

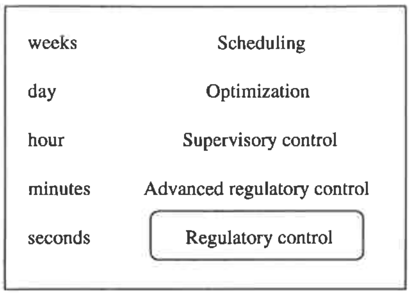


Figure 2: Control system hierarchy

forming the DOF analysis as a "tick-off" procedure identifying possible single-loop control structures (pairings). Some of these variables may be constrained, for example using the feed to a distillation column for control is limited by flooding and weeping conditions in the column. This may create a need for using more DOF's than objectives for control.

Note that the selection of overall control objectives will affect the subsequent control system design. Consider the flowsheet in figure 1 where "Unit i" can be a single or several aggregated unit operations. Here three alternative objectives determine the production rate: 1) process all available feed F, 2) maximum production due to intermediate variable I reaching constraint or 3) produce given amount P. This decision will decide if inventory control will be in the direction of flow or opposite. The best choice (given available alternatives) will also consider how integration links the process subsystems together. Direction of control should enclose subsystems generating disturbances.

3.2 "Bottom-up" design of control systems

The control system is usually designed "bottom-up" and the final control structure may of course be different from the pairings which were more or less arbitrary found in the tick-off DOF analysis.

Control systems are usually designed in a hierarchical fashion to accomplish stabilization, servo-control, optimization, fault detection and more. This gives a control system that is divided by the "time-constant" of the given operation as shown in figure 2. The regulatory control system will stabilize the plant, remove the effect of disturbances and provide "new" degrees of freedom (setpoints) for the supervisory and optimizing control to manipulate the long-term operation of the plant.

A tentative schedule for constructing control loops must first give attention to stabilizing any unstable or very slow modes including variables that may reach constraints (such as variables which are not "self-regulated" or are very sensitive to disturbances). In general one should use manipulated variables close to an objective for control, and seek to avoid using manipulated variables that may easily reach their constraint

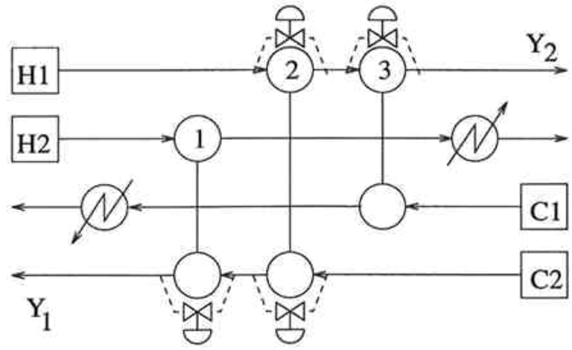


Figure 3: Heat exchanger network with alternative control bypass placements.

or impose performance limitations for higher levels (not easy). Second, one should close other important loops, these are typically quality loops (compositions). Any remaining degrees of freedom can be used for optimizing the plant either statically or dynamically.

Material balance control. It is obvious that one cannot have good flow control as well as good level control in the same process section. This may be resolved by understanding which variables are critical and which can and should be able to fluctuate. For example "sloppy" control of holdups is beneficial to reduce flow variations.

3.3 Implications for integrated plants

Degrees of freedom have so far been treated as a commodity of finite size; really the essence of the tick-off procedure. Unfortunately, this does not always hold. The last DOF and the last objective may end up at each end of the process, giving an unrealistic solution. With many actuators there is also the problem of matching the speed of response. Although the controllability analysis may point this out, high frequency disturbances or fast inventory control may render a slow actuator without the ability to comply with the needs.

The use of available DOF must also be weighed against the overall objective of the process; are there conflicts between the regulatory control and overall objectives? This will be illustrated by an example, the heat exchanger network in figure 3. The network is shown in the grid representation, with hot streams H_1 and H_2 being cooled by heating up cold streams C_1 and C_2. The heat exchangers (represented by the vertical connections) are denoted 1, 2 and 3 and there is also a heater and a cooler in the flowsheet. The control objectives are the temperatures Y_1 and Y_2 and possible bypass locations are indicated. Common rules for regulatory control of heat exchanger networks advise placing the bypasses close to the objectives, which favors using the bypasses on exchangers 1 and 3. However, exchanger 2 is an "inner match" exchanger, that is, both exit streams continue to other process stream exchangers. *Not* using exchanger 2 for control will incur an energy penalty in the heater and cooler. In this case the favorite bypasses for regulatory control are a bad choice in terms of energy usage.

4 Effects of integration

4.1 Operability limitations

The problems in controlling integrated plants come from the fact that constraints are introduced and the number of DOF's change compared to the original process. Many areas in the plant construction give rise to control problems, including:

Heat Exchanger Networks:

- There must exist separate paths for all control loops.
- The gain of the manipulated variables decreases sharply when the effect must pass through several exchangers and may yield problems with constraints.

Recycle systems:

- Effective gain increases.
- Inverse responses may occur for some control configurations using purge flowrate as manipulated variable.
- Using purge/recycle for control depends on the available gain (relative size) in the system and the effects of redirecting disturbances.

Plant-wide:

- Interconnections often remove DOF, while recycle adds a DOF. This means that experience gained from controlling single units may not apply.
- Effects of disturbances and control action propagate further, i.e. applying control may disturb other objectives (ex: bypass).

We are here always interested in improving the behavior of the closed-loop system, which is usually much faster than the corresponding open-loop system. An example of this is a distillation column, which may be very sluggish in open-loop, while the problem in closed-loop is the interaction between composition control loops, not sluggishness. In closed loop the effects of integration may be more or less pronounced than open loop.

Despite this goal, there are also benefits from investigating the open-loop characteristics of the system, one example being the frequency dependent behavior used in controller design. We also use the notion of "perfect control" in investigating the closed-loop behavior, thus avoiding to actually create a controller for the given system for analysis purposes.

4.2 Operability tradeoffs

Introducing integration may consist of many steps for a large process plant (during design or retrofit). Each step should ideally contribute to a single definable improvement to ease evaluation. This is not always the case (as with any process modification), leading to tradeoffs between plant properties. Some of the possible contradictory decisions that arise are:

Tight integration. A base case design is often performed before integration is evaluated, and during this phase the critical control objectives are usually established. The development of an integrated plant depends on all the elements/units combined, so the number and nature of objectives may change through integration, obscuring the basis for comparison.

Constraints. Applying control to account for perturbations will at the same time require that the process is operated a distance from the important constraints (Perkins, 1989). The amount of slack that separates the operating point from the constraints is a measure of how important these constraints are, a realization of how good the control is and how large the expected perturbations are. Better control will allow larger savings due to operation closer to the constraints. Integration may reduce the available gain of the actuators and demand larger back-off than the unintegrated plant.

Operability breakdown. There exists a tradeoff between the individual operability issues, each giving different limitations on the plant. "Flexibility" may demand certain equipment ratings to cover all operating points. "Controllability", considering transient behavior, may necessitate larger, smaller or reconfigured equipment. Preferred control strategy may also differ considering different timescales. Both issues will vary within the operating space and which optimal operating point to choose will also affect the mentioned criteria.

Process pinches and control. Process pinches may appear between process units or within the units themselves. Their effect on control depends on which process perturbations are to be handled and which manipulated variables are used. Heat and composition perturbations will often be damped when approaching/crossing a pinch. This is a beneficial effect if the disturbance and critical measurement are separated by a process pinch. On the other hand, the pinch may make a measurement insensitive to the intended manipulated variable.

Controlling intermediate variables. The critical measurements in a control system may themselves not be the best or only alternatives for control in a plant. Intermediate variables may present a better control solution because they are easier to control, more accessible or create an additional damping in an important process unit.

In HENs for example, pinch design and energy efficiency rules recommend having utilities only at the end of the heating or cooling procedure. This situation may change when there are process streams that are hotter than the hot utility or colder than the cold utility. Utility exchangers may then be unavailable for final temperature control, although these exchangers may dampen perturbations in the interior of the HEN.

Heat integrated networks may suffer from too few possible manipulated variables with sufficient gains for control. Thus the incentive to reduce disturbances where possible arises.

Overdesign. A flowsheet and the embedded unit operations are always developed according to some assumptions about the plant, operating conditions, proposed production and purity specifications. Since these may change over time some overdesign is always included. Additional attention to overdesign may arise from issues such as benefits in control and anticipation of increased demand, i.e. debottlenecking. The latter question is highly plant individual, while the first shows some generality. Several authors have looked into the control benefits from overdesign, for example, Jacobsen and Skogestad (1991) have investigated the effects on distillation columns. The cost/effect tradeoff of overdesign may move significantly under integration.

Variable mismatch. Mismatches may exist between manipulated and controlled variables that enable small disturbances in one stream/channel to adversely affect others. Temperature control can be seen as a mere stabilizing effect at one point of the process, while the same temperature is vital for composition disturbances in following flash tanks. Consider flashing a two-phase stream whose composition spans many components. A relatively small temperature change may shift one component from a product stream to another, thus disrupting later purification stages.

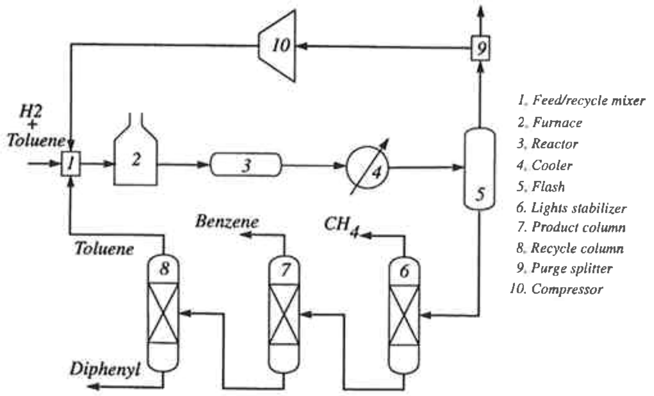
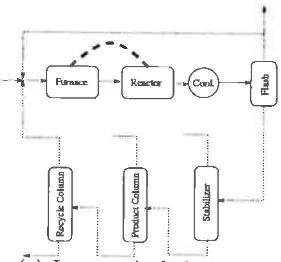
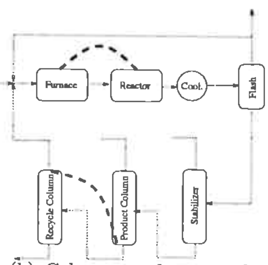


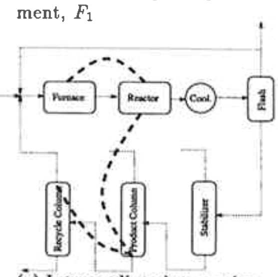
Figure 4: HDA-process with no heat-integration, flowsheet F_0



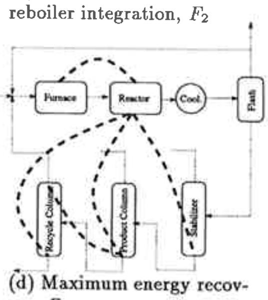
(a) Largest single improvement, F_1



(b) Column condenser and reboiler integration, F_2



(c) Intermediate integration scheme, F_4



(d) Maximum energy recovery, F_6

Figure 5: Alternative integration schemes for the HDA process

5 Illustrative cases

5.1 HDA plant

Douglas (1988) has proposed several alternative flowsheets for the conversion of Toluene to Benzene (hydrodealkylation of toluene, or HDA-process), including varying degrees of heat integration. The un-integrated alternative is shown in figure 4. Some alternative heat integrated flowsheets are shown in a condensed representation in figure 5. Here a dashed thick line means an added heat exchanger between the indicated process units; the heat transferred links two process units together. In general we therefore denote them heatlinks. Not only will the proposed integrated flowsheets have different structural paths for disturbance propagation between process units, but a parameter optimization will give each plant a different operating point. This and other features complicate the evaluation of operability between the different plant alternatives.

We will show that the operability issue for heat integrated plants may depend on the number of, functionality and placements of the heatlinks.

Following the development of a more integrated flowsheet (increase in the number of heat links in the process), the pos-

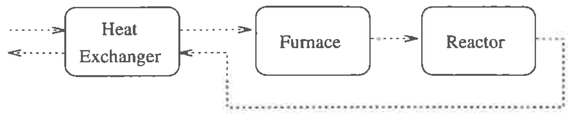


Figure 6: Feed effluent heat exchange around reactor.

Measurement	Manipulated variable
Flash inlet temperature	Pre-flash cooler duty
Production rate	Gas feed flow
Product purity	Product column reflux ratio
H_2/Aromatics ratio	Toluene feed flow
Flash outlet pressure	Purge valve opening

Table 1: Preferred pairings.

sible additions for the HDA process are:

F1. Reactor feed-effluent heat exchanger.

F2. Recycle column condenser to Product column reboiler.

F3. Reactor effluent to Stabilizer boiler, or

F4. Reactor effluent to Product column boiler.

F5. Both additions 3 and 4.

F6. Reactor effluent to Recycle column reboiler.

F_i with indices 0 to 6 represent the flowsheets mentioned above with 0 being the plant with no heat integration. The same indices are also used in comparing operability. The recycle column operates at elevated pressure in flowsheets $F_2 - F_6$ to accomplish the integration.

Heat exchange between reactor feed and effluent will generally look like figure 6, which corresponds to the F_1 case. Additional energy recovery ($F_2 - F_6$) is done by splitting this heat exchanger and letting the effluent heat other sinks intermittently.

An operability study has previously been done on the least integrated flowsheet, i.e. F_1 (Wolff *et al.* 1994), this addition being the most significant for energy savings. Control of the reactor inlet temperature was part of the design for stabilization purposes. The control loops given in Table 1 were recommended. The role of these (and other) manipulated variables must be investigated with respect to how they can be transferred to use in the more integrated cases.

Limitations and tradeoffs through integration. Earlier studies (e.g. Wolff *et al.* (1994)) have primarily considered two disturbances (downstream pressure and colling water temperature), but other scenarios are also discussed in the following.

Flowsheet F_1 involves a feed/effluent exchanger that destabilizes the reactor. Two solutions for alleviating this exist, both will also remove other temperature disturbances to the reactor. Using a bypass around the exchanger for stabilization would need a prohibitive large bypass. Stabilization through controlling the furnace exit temperature (and thus the adiabatic reactor outlet) is instead recommended from economic reasons.

F_2 demands higher pressure in the recycle column (507 vs. 101 psi) which might give added separation costs. The condenser is run utility free, so any control must involve a bypass. Slack control (for example due to constraints) seems admissible, since increased diphenyl content will move the biproduct reaction towards the intented product.

F_3 links the stabilizer reboiler and reactor effluent, limiting the control over the H_2, CH_4 content in the stabilizer bottoms, and thus the feed to the product column.

F_4 is very similar to F_3, integrating the product column reboiler with the reactor effluent, but different in that there is no integration upstream of the product column.

F_5 links the stabilizer and product column reboiler through the the reactor effluents course. This creates a new path for disturbances to propagate, especially since any control with the stabilizer boilup must be through a bypass, which will generate disturbances for the product column.

F_6, with the added integration of the reactor effluent and recycle column reboiler, has removed almost all direct control actuators from the distillation train. There seem to be few means to maintain flexibility even if the plant can be run at the optimal operating point.

Evaluation of integration schemes. To summarize; flowsheets F_1, F_2, F_3 and F_4 seem to have a good possibility of being both controllable and flexible. F_5 and F_6 create additional problems by introducing new disturbance paths. Flexibility may also be at loss with primarily bypass actuators in the separation system.

The identified disturbances are already taken care of and the recommended actuators are not modified by the integration. The interaction properties of the process, as indicated by the diagonal elements of the relative gain array

$$\Lambda_{F_1}(0) = [1.11\ 1.03\ 0.56\ 1.03\ 0.47]$$

indicate that interaction is not the most prevalent problem of design F_1. The integration schemes that are proposed lie mostly downstream of, or far away from the preferred control loop as found for flowsheet F_1. From this it is conjectured that interaction is not the primary difficulty for the proposed flowsheets.

From the discussion above, no "negative" conclusions seem possible, although a "threshold" appears to exist between F_4 and F_5 (from the introduction of new disturbance paths), with the former being more feasible.

As long as the development of the process design and the control system are sequential tasks this approach can be rewarding in eyeing integration alternatives that can cause control difficulties. A more concurrent design/control development scheme will necessarily differ. Startup and shutdown issues may also discard design schemes.

5.2 Example: conflicting objectives

This example, a column pumparound, illustrates how integration, despite creating additional degrees of freedom, displays problems due to lacking leverage in the manipulated variables.

Problem description. Refinery distillation often uses pumparounds to manipulate specific cuts from the different column sections. A simplified flowsheet of such a system is given in figure 7. Pumparounds may be quite large and utilizing the latent heat in these streams is economically significant. In this example the pumparound stream from column C_1 goes through two small heaters before heating stream F of variable size and reboiling column C_2. The first two heat exchangers may be viewed as disturbances only. HX_1 has a duty in the range $1 - 14MW$ depending on the feed to the plant. There is a deadtime of approximately three minutes between the major units in the flowsheet. The pumparound should remove a

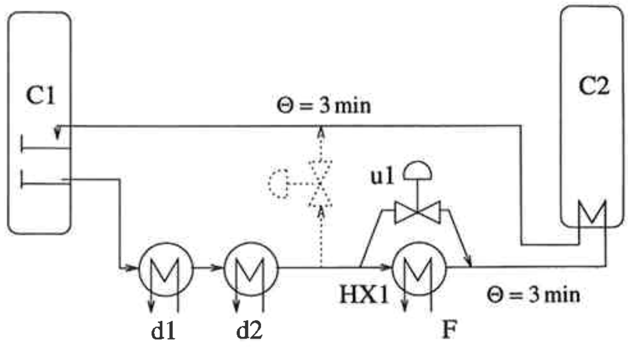


Figure 7: Distillation column pumparound

constant duty from column C_1 while providing a stable and reliable energy source for column C_2.

Limitations and tradeoffs. Problems arise from the the varying duty in the heater HX_1, since the power of the manipulated variable, the bypass fraction, is nearly diminished at the lower operating point. The desired duty in C_1, C_2 and HX_1 all put constraints on the operability. Large deadtimes between units are also detrimental. Clearly, excellent control of C_1 will have at least some interests contradictory towards control of C_2.

Better overall control can be reached by adding an auxiliary cooler or heater or redesigning the piping and control scheme. Adding control to bypass HX_1 and C_2 altogether (indicated in figure) will give faster control of the duty removed from C_1, at the cost of disruptions to the cold side of HX_1. Moving the heat loads from serial to parallel operation is also possible. Using split ratios for control instead of bypasses may give more flexibility (larger driving forces in all exchangers) and better switchability (trade effect in HX_1 with $d1, d2$). Which objective to give priority (C_1 or C_2) will also be more independent with parallel exchangers.

6 Guidelines

Interaction, constrained manipulated variables and changes in DOF limit control alternatives in integrated plants. We present the following guidelines for control of integrated plants:

- Choose preferably manipulated variables that do not disturb other process equipment.
- Make sure disturbances are not redirected or propagated to sensitive parts of the process.
- Look for control solutions that give added benefits to other objectives, for example help stabilize other equipment.
- Prefer integrating column condenser/reboiler if not important manipulated variable.
- If the condenser/reboiler is important, assure that bypass gain or auxiliary exchanger is large enough for control purposes.
- Consider shifting heat loads in parallel instead of in series for better control.
- Prefer process modifications that are beneficial to several facets of operability (flexibility, controllability, etc.).

- Integrated flowsheets should be reoptimized w.r.t. total annualized cost since energy reduction savings can be increased by transferring them to raw material savings or vice versa. An example of this is to increase the recycle, which gives better rawmaterial utilization, when associated energy usage goes down due to integration.
- Feed-effluent heat exchangers around exothermic reactors are sources of instability. A stabilizing controller can also dampen or remove other perturbation sources and thus need less control than the non-integrated flowsheet.
- Controlling pressure rather than flowrates gives better disturbance rejection properties in gas-phase inventory control. This should also create more stable operating conditions in pressure sensitive equipment (VL-equilibrium) such as columns and flash tanks. Avoid using manipulated variables that will lead to composition changes in downstream units.

7 Conclusion

It is possible to identify areas within process integration that have a large effect on the operability of the plant. The constraints imposed by integration can severely limit the operation space and be detrimental towards disturbance rejection. Feed-effluent heat exchangers remove inherent stability, but this induces no loss in operability.

We have proposed guidelines for the development of operable plants in view of several integration topics. Integrated plants exhibit problems extending the scope of plant wide control.

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