Towards Integration of Controllability into Plant Design

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Abstract

In process design practice the plant piping and instrumentation diagram evolves iteratively using mainly experience and process reasoning to address questions related to plant controllability. It would be desirable to be able to address such questions more quantitatively at different abstraction levels during process design such that controllability evaluation can be integrated into the design process. Only few attempts have been reported towards integrating controllability investigations into early stages of plant design. This paper reviews literature on controllability from the perspective of controllability assessment with the aim of identifying tests, which may be used at different stages of plant design. First definitions of terms used within process flexibility design and controllability assessment for control structure design are given. Methods for controllability evaluation are reviewed for some modes of process operation. Basically two types of evaluation and design methods prevail. One type is based on linear model analysis, whereas another type is based on physical chemical insight and thus provides nonlinear information.

Control structure development for controllability is illustrated on an energy integrated distillation plant by using a heuristic process knowledge based method to develop a basic control structure and subsequently using an optimisation based approach for selecting a product purity control structure. Controllability properties of a more energy efficient process design alternative is discussed to illustrate the potential trade off by choice of the most energy efficient design that however has the lowest controllability. Based on the review and the examples a procedure for integrating controllability assessment for control structure development into plant design is proposed.

Introduction

From the introduction of the first version of general purpose process controllers during the first quarter of this century, control has developed to become an indispensable part of process operation to stay at the competitive edge for many products, to satisfy product quality and often also satisfy environmental requirements. This trend develops even further as the plant heat and mass flows become more tightly integrated and as plants are optimised both in design and operation such that nonlinear features are exploited to a higher extent. The gradual development and incorporation of control methodologies have probably contributed to the present state of ad hoc design habits, where the P&I-diagram at best is developed iteratively during the plant design. In this type of procedure little regard seems to be paid to plant controllability aspects even though the reason for control is precisely that of ensuring that the plant is controllable subject to known

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demand and disturbance variations and also subject to all uncertainties imposed during plant design.

Plant design evolves as a sequence of decisions and evaluations. Traditionally one of the early plant design decisions is to select the pertinent process operation mode, i.e. continuous, periodic or batch. This operation form issue should ideally be settled fairly early during the design procedure as the choice has important ramifications on several steps in the design procedure. The periodic and batch operation modes are dynamic, hence their design procedure is somewhat different and especially different from the static design procedure used for traditional design of continuous plants. Similar differences are relevant for controllability. Thus the review will also address differences between controllability of these operating modes. Note that start-up and shut-down may be considered as special cases of fed-batch or batch operation. Controllability assessment precedes the control system design during the plant design procedure. Thus controllability assessment deals with whether the plant is controllable, and ideally also with what is the achievable controlled performance of the plant. How the plant actually is controlled is subsequently dealt with during pairing of measurements and actuators, i.e. control structuring and during controller design.

The purpose of this paper is firstly to review concepts of controllability in relation to similar concepts. Secondly methodologies for assessing controllability are reviewed from the perspective of identifying tests, which may be used at different stages of plant design. Thirdly control structure development is illustrated on case study to reveal the usage of different tests. The paper is organized as follows: Methods for controllability evaluation are reviewed for different modes of process operation. Control structure development for operability is illustrated on an energy integrated distillation plant, using two different methods. The trade off between optimal process design and operability is also illustrated. Finally a procedure for integrating controllability assessment for control structure development into plant design is proposed.

Basic Concepts

Traditionally plant design has been a mostly sequential discipline, where the control design is carried out after the plant has been designed. Today plant design is viewed as an iterative procedure where the P&I-diagram also is developed iteratively. Controllability evaluation is intended to be relevant also at earlier design stages than where the P&I-diagram is used today. During plant design a number of basic plant performance requirements have to be ensured in order to obtain a design which provides acceptable operational performance.

Operability is the ability of the plant to provide acceptable static and dynamic operational performance. Operability includes flexibility, switchability, controllability and several other issues.

Flexibility is the ability to obtain feasible steady state operation at a number of given operating points, i.e. over a range of uncertain conditions. These uncertain conditions can be defined from expected variations in raw material and in process performance.

Switchability is the ability to switch between operating points. The main issues are dynamic feasibility and safety. For service plants fast switching may be desirable to minimise loss of product and energy consumption.

With the above definitions the commonly used term feasibility has both a static aspect, which is incorporated into flexibility and a dynamic aspect, which is part of switchability. Since the methods for flexibility evaluation have developed also into dynamic flexibility it seems relevant also to consider their relations to controllability.

1 Controllability

Controllability is used with several different meanings in literature. A few of these are discussed below with emphasis on the way controllability may be used during process design.
Two main approaches have been the basis for definition of controllability. One is based on a goal or purpose oriented view while the other is based on a mathematical or state space oriented view. Some of the goal oriented definitions are discussed below and complemented with a basic mathematically oriented definition to indicate how a practically useful definition has been reached.

In dealing with continuously operated processes Ziegler and Nichols (1943) defined controllability as:

> *The ability of the process to achieve and maintain the desired equilibrium value*

Rosenbrock (1970) defined controllability in more general terms:

> *A system is called controllable if it is possible to achieve the specified aims of control, whatever these may be. By extension, the system is said to be more or less controllable according to the ease or difficulty of exerting control.*

Thus controllability may be viewed as a property of the plant, which indicates how easy it is to control the plant to achieve the desired performance. Rosenbrock (1970) introduced the term functional controllability:

> *The system is functionally controllable if given any suitable vector y of output functions defined for $t > 0$, there exists a vector u of inputs defined for $t > 0$, which generates the output vector y from the initial condition $x(0) = 0$*

Functional controllability, however, only provides a yes/no type answer, and gives no measure of achievable performance in case the process is not functionally controllable. Dynamic resilience was introduced by Morari (1983) as

> *The quality of the regulatory and servo behaviour, which can be obtained by feedback.*

Thus this concept is closely related to functional controllability, but dynamic resilience also include a quality measure of the achievable performance independent of the controller.

In control theory literature the term controllability has only little connection to the ease with which a plant can be controlled. The following definition is termed 'state controllability':

> *A state is termed controllable if for any initial state $x(0) = x_0$, any time t_1 and final state x_1, there exists an input $u(t)$ such that $x(t_1) = x_1$.*

In control theory literature a system is termed controllable if all states of the system are 'state controllable'.

Analogously 'state observability' is defined:

> *A state $x(t)$ of a process is termed observable at some given t if knowledge of the input $u(t')$ and $y(t')$ over a finite time $t_0 < t' < t$ completely determines $x(t)$.*

A process is completely observable if all states are state observable. The above definitions of controllability and observability are often referred to as 'Kalman controllability and observability' due to their inventor Kalman (1960). Kalman also showed that a linear model may be decomposed into its controllable/uncontrollable and observable/unobservable parts using a similarity transformation.

In practice however it may be difficult to obtain acceptable control of all states in a process, even if all states are controllable. Any unstable state must be both state controllable and state observable, in order to close feedback paths around it and thereby stabilise it. Thus the following concepts become useful:

Stabilisability: A process is stabilisable if there exists a controller K which can stabilise all unstable modes. In the linear case this requires that $A - BK$ is stable, i.e. has all its eigenvalues within the left half plane.

Detectability: (Linear case) A process is detectable if the unobservable subspace does not contain any unstable modes. Thus an observer may be constructed for the unstable modes. Thus it is only necessary to require that the unstable modes are stabilisable and observable. Based on this requirement **input-output controllability** may be defined (Skogestad, 1994; Skogestad and Postlethwaite, 1996):

Input-output controllability: The ability to achieve acceptable control performance,i.e. to keep the controlled outputs (y) and manipulated inputs (u) within specified bounds from their setpoints (r), in spite of signal uncertainty (disturbances (d) or noise (n)) and model uncertainty, using available inputs and available measurements.

With input-output controllability as definition for controllability, this concept is a plant property which reflects how easy it is to control the plant. Plant controllability depends on many different aspects such as specific plant dynamics, sensitivity to uncertainty, measurement location, actuator constraints and disturbance characteristics. These aspects are discussed below in a brief review of methods for controllability evaluation.

2 Controllability Evaluation

Controllability may be investigated using mainly three approaches,

1. A process understanding (i.e. thermodynamics) oriented approach
2. An optimisation approach
3. A non-linear, e.g. a passivity approach

The first two approaches may be based on linear models where these apply, whereas the latter approach, which is not yet fully developed, applies for nonlinear models. These approaches are reviewed below to provide a basis for discussion of approaches for including controllability analysis into process design at an early stage. To partially follow a historical perspective the linearised models are treated first. The focus is on conveying the basic ideas, advantages and limitations of the methods rather than details which may be found in the referenced literature.

2.1 Evaluation for Linear Models

A wealth of controllability measures, which are based on a linear model or a transfer function for the plant have been proposed. These measures may be viewed as analytical or process oriented in that they provide a fundamental understanding of what limits controllability, i.e. the achievable performance of the controlled plant. The process is modelled as: $y(s) = G(s)u(s) + G_d(s)d(s)$, where y is the measured outputs, u the manipulated inputs, d the disturbances while G and G_d are the plant and disturbance transfer functions respectively. The reference inputs are y_{ref}. The control error is $e = y - y_{ref}$. It is assumed that all variables are scaled to be within -1 to 1 by dividing the unscaled signals with their unscaled maximum expected allowed change, Skogestad and Postlethwaite (1996). The conventional feedback controlled process with controller transfer function K is shown in Figure 1

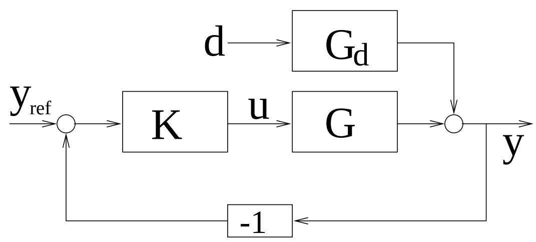

Figure 1. Conventional feedback blockdiagram for a process with conventional controller (K)

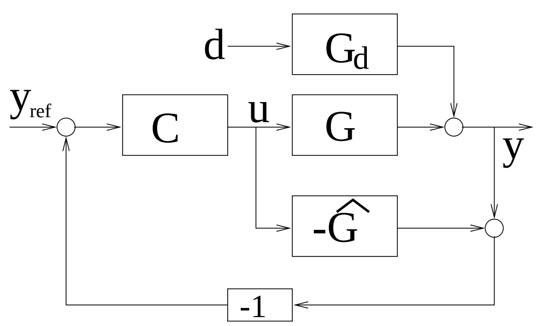

Figure 2. Internal Model Control feedback blockdiagram with internal model controller (Q)

Plant characteristics which limit achievable performance of feedback control may be outlined based on the concept of *internal model control* (IMC) (Holt and Morari, 1985a and b; Morari and Zafiriou, 1989). In IMC a model is used in parallel with the plant, as shown in Figure 2, such that the feedback signal is the difference between the two outputs. Thus the feedback signal only contains model plant differences, unmeasured disturbances and noise. Hence if a perfect plant model and perfect knowledge of disturbances are available then no feedback is necessary, i.e. control can be entirely feedforward based. The transfer function from setpoint to output is $y = GQy_{ref}$, therefore nominal stability is only guaranteed if both the plant and the controller are stable. The intuitively appealing design of an IMC regulator becomes obvious in that perfect control is achieved for $Q = G^{-1}$. Hence any limitation on constructing the plant inverse is a cause for imperfect control, and therefore constitutes a limitation on achievable control performance. Several limiting phenomena may be listed.

1. Time delays, which when attempted inverted become predictive, which cannot be perfectly accomplished in a controller which does not know which inputs may arise in the future.

2. Plant zeros in the right half of the complex plane (RHP), which become unstable poles in the plant inverse and therefore render the ideal controller unstable.

3. Unstable poles requires control. To apply IMC design special precautions must be taken (Morari and Zafiriou, 1989). Simultaneous presence of both RHP pole and zero requires that the zero is further away from the imaginary axis than the pole for design of a stable controller.

4. Pole excess of the plant transfer function will render the perfect controller improper. A proper controller is obtained by also using a suitable order low pass filter in the controller.

5. Constraints in the manipulated variables. The magnitudes of the disturbances which can be rejected are limited by the actual constraints on the manipulated variables. The static issue here is that of flexibility, but the general problem depends on the disturbance frequencies. As the controller gain in conventional control approaches infinity to achieve perfect control the control error $e = y - y_r \rightarrow 0$ thus $u = G^{-1}y_r + G^{-1}G_d$. Hence if $|[G^{-1}G_d]_{ij}| > 1$, then disturbance j can cause imperfect control of variable i. A plot of the frequency function of the elements of $[G^{-1}G_d]$ provides insight into the possibility of violating input constraints.

6. Model uncertainty will also limit achievable performance. In multivariable plants inversion of the plant transfer matrix becomes difficult as this matrix approaches singularity. Hence the plant inverse should not be used directly for control. Singularity may be evaluated using a singular value decomposition: $G(i\omega) = U(i\omega)\Sigma(i\omega)V^H(i\omega)$, where the singular values are contained in the diagonal singular value matrix: $\Sigma(i\omega)$. The condition number is the ratio of the largest and smallest singular values: $\gamma(G) = \frac{\bar{\sigma}(G)}{\underline{\sigma}(G)}$ and depends on scaling. To avoid ambiguity a minimised condition number γ^* is used, where G is pre- and postmultiplied by real diagonal scaling matrices: $\gamma^* = \min_{D_1,D_2}\gamma(D_1GD_2)$. A large minimised condition number indicates an ill-conditioned plant.
   An interaction measure is the relative gain array (RGA): $\Lambda(G(s)) = G(s) \times G(s)^{-1}$ where $\times$ denotes element by element multiplication. The ij'th element of Λ can be shown to be the ratio of the open loop gain from input j to output i when all other loops are also open, to the gain from input j to output i when all other loops are perfectly controlled. The RGA is very useful as it is scale independent. A relation between the minimised condition number and RGA has been established by Nett and Manousiouthakis (1987): $2\max||\Lambda(G((i\omega)||_1, G((i\omega)||_{\inf})) \leq \gamma^*(G(i\omega)) + \frac{1}{\gamma^*(G(i\omega))}$.
   Yu and Luyben (1987) proved that if a single element of G is perturbed from g_{ij} to $g_{P_{ij}} = g_{ij}(1-1/\lambda_{ij})$ then the perturbed matrix G_P becomes singular. Thus if an individual element in the plant transfer function has an uncertainty larger than $|1/\lambda_{ij}|$ then the plant may have RHP zeros at the frequency where this occurs. Thus large relative gain elements result in extreme sensitivity to uncertainty. This result has implications for both identification and control, in that plants with large RGA elements will be difficult to identify and also difficult to control due to large sensitivity to model uncertainty.

Each measure described above treats one of the control performance limitations and provides information on the qualitative performance limitations but does not relate directly to the performance requirements.

The main limitations of the linear analysis techniques are that they are based on input-output models and it may be difficult to relate control performance limitations directly to design variables. Another limitation is the evaluation of influence of model uncertainties which is directly related to the specific uncertainty description. Most of the linear analysis tools are based on frequency domain specifications, whereas often time domain performance specifications are desirable. Finally the application of linear controllability methods require considerable experience as each indicator usually only considers one of the control performance limitations. Thus there still is a need to further develop even the linear analysis tools and also develop methodologies which further their usage within a process design context.

2.2 Optimisation Methods

Controllability may be evaluated by using an optimisation formulation. If the model is assumed linear then many of the above measures may be imposed as constraints to the optimisation problem. These methods will be reviewed under nonlinear techniques, due to the possibility for simple extension to nonlinear models, where however convergence proofs are limited.

3 Controllability for Nonlinear Plants

The controllability definition adopted in this paper is based on the goal or purpose of control, hence this definition applies equally well for nonlinear plants. Nonlinear plants however may exhibit much more complicated dynamic behaviour than linear plants, thus methods for evaluation of controllability may be somewhat different. Even though most plants are truly nonlinear, linear approximations may describe plant behaviour within parts of the operating window. A key property which exemplify nonlinear behaviour is the occurrence of a characteristic change of behaviour within the operating window. Such a change can be the occurrence of multiple steady states for some range of operating parameters, where each of the steady states will have individual stability properties. Another characteristic change of behaviour occurs when a previously stable steady state turns unstable. Such behaviours have been described in several cases for both reactors and separation processes. The operating parameter values at which the characteristic changes in behaviour occur are called bifurcation points. When considering controllability it is of course important to know about such points. One may consider operating the plant outside the region with complicated behaviour. It appears however as if attempting optimal design and operation exploits nonlinear behaviours of process plants (Jørgensen and Jørgensen, 1998) such that bifurcations occur somewhere around the optimal operating point. However overdesign may reduce the tendency to complex behaviour (Seader *et al.*, 1990), but clearly an integrated approach to plant design and operation optimisation could be most competitive. In such an integrated approach a first step would be to perform a bifurcation analysis to reveal the types and locations of possible bifurcations. Subsequently each type of behaviour can be further analysed also using linear methods. Thus the controllability analysis may be treated through a number of local analyses.

Methods for controllability analysis of nonlinear plants are far less developed than for linear plants. Some aspects of analysing controllability of nonlinear plants are given below, where two types of nonlinear controllability measures which provide some insight are described. Finally optimisation methods are described.

3.1 Analytical methods

In nonlinear dynamics a nonlinear inverse may be evaluated. Since there are no direct method for quantifying the effect of inverse dynamics, the nonlinear inverse must be analysed instead. One approach to directly addressing the question of whether the inverse of a dynamic system is stable is to analyse the inverse dynamics, also called *zero dynamics*. The zero dynamics is given by the dynamics of a minimal order realisation of the system inverse. The analysis may be grouped into two cases:

1. For constant setpoint the stability of the closed loop with the (right) inverse employed as a controller is completely determined by the stability of the unforced zero dynamics. Daoutidis and Kravaris (1991) describe a nonlinear system as minimum phase if its (unforced) zero dynamics is asymptotically stable, and nonminimum phase if it is unstable.

2. For a reference trajectory tracking, i.e. a servo problem, the inverse dynamics is driven by the desired system output trajectory and its first $r-1$ derivatives, where r is the relative degree, which characterises the lowest order derivative of the output y that is explicitly dependent on u. In this case the forced zero dynamics must be evaluated to determine internal stability.

Similarly to the usage of RGA for linear systems, the static RGA may be used as a measure of the effect of uncertainty on controllability (Mijares *et al.*, 1985). The block relative gain (BRG) is extended to nonlinear systems by Manousiouthakis and Nikolaou (1989) to provide a static NBRG and a dynamic version (DNBRG). The static NBRG is shown to be a lower bound for the condition number for the nonlinear system. However it is not clear how these two measures relate to achievable control performance.

3.2 A Passivity based Methodology

An interesting new development is passivity based control (Farshman, Viswanath and Ydstie, 1998). The merits of this promising methodology are that it is possible to synthesize a guaranteed stabilising control configuration based on model information only, where each inventory is controlled. The underlying control design can be relatively simple, such as multiple proportional and integral regulators (inter-)connected with a number of additional measurements to ensure feedforward knowledge about the loads on the different loops. Two disadvantages are that this far only relatively few processes have been formulated into model representations which fit into this framework and that the methodology at present requires measurements or estimates of all state variables. This latter requirement is the case for many nonlinear control techniques. A key property of passive systems is that subject to some smoothness requirements a system constructed by an arbitrary interconnection of passive systems is itself a passive system. Controllability aspects of this methodology are straightforward to ensure in that it is just a matter of establishing passivity for the system at hand. Thus there is significant interest in finding representations that render a system passive.

3.3 Optimisation Methods

These methods constitute perhaps the most successful methods in attempting to integrate design and controllability. These methods are reviewed by Walsh and Perkins (1996). Here some of the main developments are emphasized to illustrate assessment of controllability. The determination of flexibility is also considered since that methodology has affected assessment of dynamic flexibility, which is related to controllability.

Narraway, Perkins and Barton (1991) provided a measure of the best achievable economic performance as the amount that the operating point must be backed off from the optimal operating point to ensure that none of the operating constraints are violated thereby affecting controllability. Walsh and Perkins (1992) provided an optimistic bound on disturbance rejection performance by assessing performance of an idealised controller under worst case conditions. The plant performance was limited by both delays and uncertainty. White *et al.* (1994) evaluated switchability of a proposed design, provided the control system is given. Mohideen *et al.* (1996) extended this to consider operability analysis, control structure selection and controller tuning. Vu, Bahri and Romagnoli (1997) incorporated also operability into the switchability problem. Soroush and Kravaris (1993 a and b) addressed flexibility of operation of batch reactors. They defined flexibility qualitatively as the ability of a reactor to operate according to a predetermined optimal trajectory in the presence of uncertainty. Kuhlmann *et al.* (1998) presents an approach for design of robust controllers, which is integrated into a fed-batch design problem formulation. Chenery and Walsh (1998) proposed a linear performance measure to determine a measure of controllability:

$$\min_{K,u_0} J(K,u_0) \qquad \text{s.t.} \quad c(K,u_0,w) \leq 0 \; \forall \, w \in W$$

Thus a controller K and a reference operating point u_0 are selected to minimise the objective J while ensuring feasibility for all disturbances w within a bounded set W and satisfying the constraints c. The linear problem is formulated with a linear objective, linear model, linear constraints and LTI controller from the set of stabilising controllers. Feasibility aspects of the approach are demonstrated on an industrial case study.

3.4 Static Flexibility

Design of static processes may be described by the problem: $c(d,z,p) \leq 0$, where $p_- \leq p \leq p_+$ and $z_- \leq z \leq z_+$ where c represents the process equality and inequality constraints, which are indexed with i. d is the vector of design variables, z the vector of control actuator variables and p the vector of uncertain parameters. Grossman *et al.* (1983) showed that the feasibility problem for processes with the above description is equivalent to the following optimisation problem:

$$J^F(d) = \max_p \min_z \max_i f_i(d,z,p)$$

Where $f_i(d,z,p)$ is the description of the process and operational constraints. If $J^F(d) \leq 0$ then the design is feasible. If however $J^F(d) > 0$ then the solution will provide a critical point p^c where the largest violation of the constraints occurs. Swaney and Grossman (1985) extended this work and proposed a flexibility index on top of the feasibility problem with $p^N - \delta\Delta p_- \leq p \leq p^N - \delta\Delta p_+$. The flexibility index $F = max\,\delta$ quantifies the ability of the process to operate at other than the nominal operating point. They also showed that under certain convexity assumptions critical points that limit feasibility or flexibility lie on the vertices of the uncertainty space. Saboo *et al.* (1985) also formulated optimisation problems to determine static feasibility and flexibility. Grossman and Floudas (1987) exploited the fact that sets of active constraints are limiting design flexibility in their mixed integer linear/nonlinear programming problem. Pistikopoulos and Mazzuchi (1990) introduced a stochastic flexibility index for processes with stochastic parameters. Grossmann and Straub (1991) pointed out that the above two step procedure establishes the flexibility analysis problem:

1. *The feasibility problem:* Determines if a given design can feasibly operate over the considered range of uncertainty
2. *The flexibility index problem:* Evaluates a measure to quantify the ability to operate in the presence of uncertainty. Above this measure is establishment of the maximum parameter range over which the design can operate feasibly.

3.5 Dynamic Flexibility

Grossman and Morari (1983) pointed out that several dynamic situations required consideration of process dynamics in flexibility analysis. Dimitriadis and Pistikopoulos (1995) extended the approach of Swaney and Grossman (1985) to dynamic flexibility, where they include time-varying uncertain parameters.

3.6 Dynamic Flexibility versus Controllability

The static flexibility problem mainly considers the feasibility issue and an index for flexibility. These aspects are also the outset for the dynamic flexibility problem of Dimitriadis and Pistikopoulos (1995), where the key point is to be able to account for parameter variations during operation. Given a solution to the dynamic flexibility issues, then the more strict requirements of achievable control performance may be addressed given solutions to the dynamic flexibility problem. Thus in general the set of plants, which are statically flexible, includes the set of plants which are dynamically flexible. This set again includes plants which may switch between a number of operating points, and that set further include plants which are input-output controllable around a given operating point.

4 Integration of Controllability into Process Design

Several methods for integrating controllability evaluation methods during the different stages of process design have been proposed. One methodology has been to use open loop indicators following the ideas of Morari (1983), and Hovd and Skogestad (1996). This approach was taken, e.g. by Luyben and Floudas (1994), who uses these measures inside a multi objective optimisation formulation. A simpler approach has been taken, e.g. by Weitz and Lewin (1996) who attempt to develop a linear dynamic model from static flowsheet simulation information and simple assumptions concerning process dynamics. This approach is extended in Gani *et al.* (1997).
The methods for controllability evaluation are briefly summarised, before a four step process design procedure is given and introduction of controllability measures at each step is briefly discussed.

1. Linear model analysis based methods. These methodologies provide a sound basis for understanding the control problems, but do not relate directly to performance requirements.
2. Optimisation based methods. A key issue is to propose a superstructure for the problem which is sufficiently rich to ensure that the truly optimal solution, in fact is included.
3. Thermodynamically nonlinear model based methods. Passivity based control evaluation may be developed to be used at a relatively early design stage to evaluate whether a candidate flowsheet can achieve a required performance from available measurements.

Below the two first types of methods are applied to a case study to develop a control structure.

5 Example

Application of controllability analysis for control structuring have been exemplified on a number of examples. One study object is the energy integrated distillation column at the Technical University of Denmark (DTU) suggested as a benchmark by Koggersbøl and Jørgensen (1995a), where both a simulation and the plant are available for further studies and experiments. Control configuration on this plant is not simple due to the tight energy integration. Below the experimental plant is briefly presented, and then some aspects of control structuring are discussed first using heuristic process understanding for developing the lower control levels on the existing plant (Koggersbøl *et al.*, 1996). Thereafter operability implications of improving the energy efficiency of the design are illustrated again using heuristic arguments. Subsequently a double purity control structure is developed through an optimisation approach (Hansen *et al.*, 1998). The optimisation based methodology provides control structures which are demonstrated to satisfy the requirements set forth in the benchmark. Finally the presented case study is used to discuss requirements for enabling evaluation of controllability at different stages of plant design.

5.1 Experimental Plant

Figure 3 shows the flowsheet of the base case plant. The column has 19 sieve trays, a reboiler, a total condenser and a reflux drum. The heat pump expansion valve (Exp. valve) throttles high pressure liquid refrigerant to the heat pump low pressure (P_L) suitable for evaporation in the condenser. The control valve (CV9) manipulates the refrigerant vapour flow rate. After super heating the vapour the compressor elevates the pressure to (P_H) suitable for condensation in the reboiler. In the base case a small part of the condensation takes place in a secondary condenser which by a cooling water circuit is connected to a set of air coolers. The cooling rate can be manipulated by the control valve CV8. Through a storage tank (Rec) and a heat exchanger the refrigerant cycle is closed at the expansion valve.
The process is modeled by a DAE model derived from energy and material balances. The model is described in Koggersbøl and Jørgensen (1995b). The simulation model includes PID-controllers for the levels in the reboiler, the condenser and the accumulator, tuned by the IMC rules described by Chien and Fruehauf (1990) to closed loop time constants of approximately 1 minute. The separation investigated in the simulation is a nearly binary feed containing 49.5 mole-% isopropanol, 49.5 mole-% methanol and 1 mole-% water impurity.

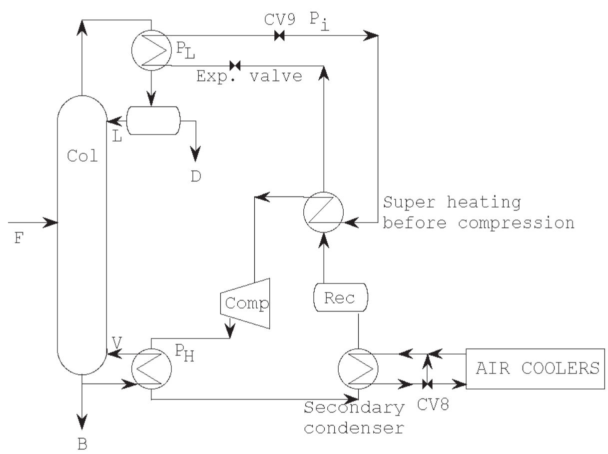

Figure 3. Schematic flowsheet of the energy integrated distillation column

5.2 Heuristic Base Level Control Structure

A heuristic control structure is developed here for the lower control levels by first considering the stability of the plant. Thereafter switchability considerations are used to select a basic control structure for this energy integrated plant. Finally the composition control loops may be implemented, the latter will be considered in connection with an optimisation based control structure. Here however the basic control structure is of most concern, since that renders this plant different from a conventional distillation column.

5.2.1 Stability

Without any control at all the plant is unstable (Koggersbøl *et al.*, 1996b). In fact with this energy recycle the open loop plant contains three unstable poles. A coupling between the accumulator and reboiler levels through the heat pump presumably gives rise to a complex pole pair. To stabilise the system one needs to control both levels and also to break the positive energy feedback in the heat pump. The reboiler level is controlled by the bottom product flow rate, the accumulator level is controlled by the reflux flow rate. All these loops are implemented by PID's. Using the reflux flow rate for level control leaves the distillate product flow for product purity control, this will result in a (D,V)-type control configuration. In Koggersbøl *et al.* (1996b) it has been experimentally verified that a (L,V)-type configuration may render the system unstable, as predicted by Jacobsen and Skogestad (1991). The remaining unstable real pole is stabilised next. There are three pressures in the heat pump which can be stabilised by manipulating either the cooling valve CV8 or the throttle valve CV9. Since CV8 directly manipulates the cooling rate (Q_{cool}), a control loop using this valve as actuator can stabilise the base case plant. In principle any of the low pressures and the high pressure could be paired with CV8 to stabilise the system. However, the gains from CV8 to the low pressures are relatively small hence this valve should preferably be paired with the measurement of P_H. This selection leaves the throttle valve CV9 free. In the reboiler the saturation pressure P_H is a sufficient measure of the condition on the refrigerant side for heat transport into the column. This condition is controlled using CV8. In the condenser which is the other contact point between the heat pump and the column the saturation pressure P_L is a sufficient measure of the condition on the refrigerant side for heat transport from the column. This condition could conveniently be controlled by manipulating the throttle valve CV9. Since the plant is already stabilised by the CV8–P_H loop CV9 may be used for control of P_L. This loop will allow column disturbances to be transmitted to P_i, through the compressors and annihilated by the CV8-P_H loop.
By controlling the conditions on the refrigerant side of the two contact points using the CV8 and CV9 actuators it thus seems possible to manipulate the heat balance of the column (Q_B and Q_C), hence the vapour flow rate and the pressure in the distillation column can be manipulated using either P_H and P_L measurements directly or using $P_H - P_L$ and $P_H + P_L$ respectively. These actuators along with the level control loops for the reboiler and accumulator, constitutes a lowest level of four control loops of which three are necessary for stable operation of the energy integrated distillation plant. It remains however to be investigated if these actuators can cover the desired operating region. This aspect is treated by considering switchability.

5.2.2 Switchability

In this section the next level of actuator configuration is addressed. The lowest or base level ensured stability. Using the setpoints of the low level controllers as the new set of actuators, switchability of the plant is analysed. It is of great importance to know if the base level control configuration ensures that the operating region can be spanned by the secondary actuators. The aim of this section is to discuss to what extent a set of secondary actuators may be used to control the energy integrated column as a conventional distillation column, i.e. specifying the conventional actuators shown as primary actuators in table 1. The latter actuators are often considered in the distillation column control literature for binary mixtures. Column pressure (P) is either assumed to be constant or controlled by the condenser cooling rate. The flow rates D, L and B are often manipulated on a molar basis which is never the case in practice, as discussed by Jacobsen and Skogestad (1991). The energy integrated distillation plant has the primary actuators shown in table 1 as discussed above.

A			
Column	Primary	Secondary	Tertiary
Top	D		
	L		
Heat	P $(P_H + P_L)$	P_L	CV9
pump:	V $(P_H - P_L)$	P_H	CV8
Bottom	B		

B			
Column	Primary	Secondary	Tertiary
Top	D		
	L		
Heat	P $(P_H + P_L)$	P_L	CVi
pump:	V $(P_H - P_L)$	P_H	CV9
Bottom	B		

Table 1. **A**: Actuator levels for base case energy integrated distillation column **B**: Actuator levels for optimally energy integrated distillation column

A set of correlations connect the heat pump pressures P_H and P_L to the conventional distillation column actuator V and the column pressure:

$$V \cdot \rho_l \cdot \Delta H^{vap} = AU_B \cdot (T^{sat}_{refr}(P_H) - T_B) = AU_C \cdot (T_C - T^{sat}_{refr}(P_L)) \quad (1)$$

$$P_C = g_1(T_C) \quad (2)$$

$$P_B = g_2(T_B) = P_C + \Delta P_{trays}(V) \quad (3)$$

where $\Delta P_{trays}(V)$ is the pressure drop across the column trays being a function mainly of the vapour flow rate. AU_B and AU_C are coefficients containing information about heat transfer areas and heat transfer coefficients for the reboiler and the condenser. $T^{sat}_{refr}(P_j))$ is the temperature of saturated refrigerant at the given pressure, and T_B and T_C are the temperatures on the column side of the reboiler and the condenser. Equations 1–3 can be interpreted as shown in figure 4-A. When P_H and P_L are brought to specified values by the CV8– and CV9–controllers the system settles at values for boil–up and column pressure(s) such that the column "balances" between the two heat pump pressures as illustrated in the mechanical analogue in figure 4-A. The vertical position of the column represents the column pressure (measured by the equivalent saturation temperature for the pure products), while the vertical column length represents the pressure drop from bottom to top which also indicates the size of the vapour flow rate. The length and the position will settle such that the temperature differences ΔT_B and ΔT_C each multiplied by the relevant heat transfer function AU_B and AU_C are equal. This way the balance (eqns. 1–3) determines the boil–up rate and column pressure, symbolised by the length and vertical position of "the column" in figure 4-A.
The gains for positive and negative changes in the heat pump pressures are illustrated in figure 4-B. Nonlinearities in pressure drop correlations and heat transfer coefficients may cause the three elastic subsystems to "deform" nonlinearly in non-constant ratios. These nonlinearities of the gains are vaguely seen on the figure even for this narrow range, as e.g. the full curve between points (1156,477) and (1156,527) is not a straight line. From this understanding it is clear that to increase column pressure at constant boil–up rate one must increase both actuators, and if

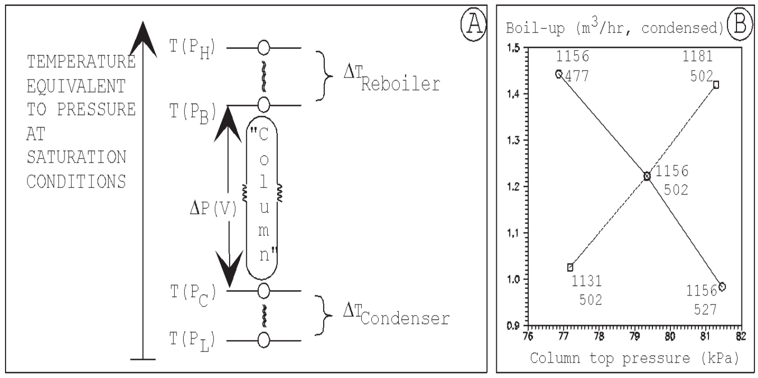

Figure 4. A: The distillation column modelled as an elastic body "balanced" (in "springs") between the temperatures corresponding to the actuators P_H and P_L. The "springs" inside the column symbolise that the pressure drop from bottom to top can vary as a function of vapour flowrate. B: Steps ±25 kPa in high and low heat pump pressures (simulation) (on plot: P_H over P_L in kPa)

the boil–up rate is to be increased at constant column pressure (either P_B or P_C) one must increase P_H and/or reduce P_L such that the column profile is "stretched" while one of the end points is maintained at the same pressure. Thus it is clear that specifying the two heat pump pressures P_H and P_L is equivalent to specifying boil–up rate and column pressure, and hence it should be possible to configure a control system manipulating the setpoints to the high and low pressures at the secondary level which in turn then are setpoints for the control valves CV8 and CV9 at the tertiary level. With this three level control structure the operator can manoeuvre the process through the operating region specifying the conventional primary actuators P and V of a distillation column thus giving the three actuator level hierarchy shown in table 1-A.

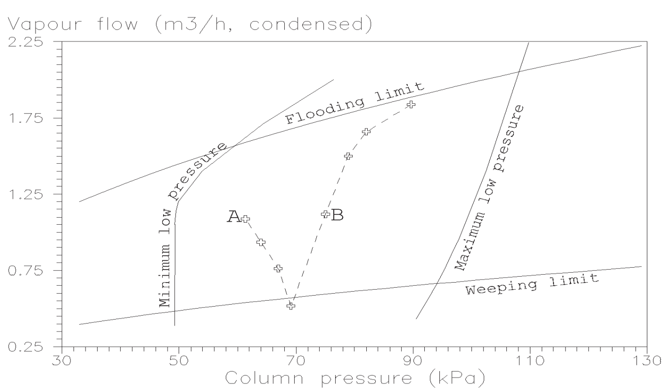

Figure 5. Experimental gain determination from heat pump pressures to top column pressure and boil–up rate. Branch A: Constant P_H, 25 kPa steps in P_L. Branch B: Constant P_L, 75 kPa steps in P_H.

Heat Pump		Q_{hot}	Q_{cold}	W_{comp}
None	SFP	288.8	288.0	0
Base case	SFP	36.2	114.8	79.4
Optimal	SFP	36.2	90.7	55.3

Table 2.: Column utility requirements for different heat pump configurations. The mechanical compressor efficiency is 76%. SFP is Separate Feed Preheater. The experimental value for the base case design heat pump COP = 4.57 is assumed valid for the other cases also. All power units are kW.

However, due to nonlinearities the feasibility of this configuration should be investigated, by determining the gains from high and low heat pump pressure to boil–up rate and column pressure over the operating region. Figure 5 shows experimentally obtained results for the variations in these gains. It is seen that as boil–up rate is decreased the gains from either of the heat pump pressures to the column pressure decrease significantly. It is clear that the suggested control configuration enabling the operator to specify P and V can suffer seriously from the nonlinearities shown in figure 3 unless the control algorithms are carefully selected and designed especially in the lower boil-up rate part of the operating region.

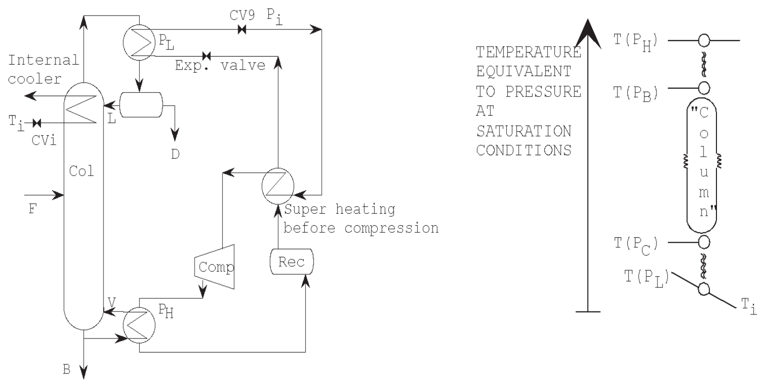

Figure 6. Optimal cooler location. The compressor heat is removed from the column top. The column now "balances" between $T(P_H)$ and a combination of $T(P_L)$ and T_i.

In the previous sections stability, actuator configuration, and switchability were discussed for the base case heat pump design. This design involves removal of the compression work on the coolant gas by cooling in a secondary condenser in the high pressure, and thus high temperature, part of the system. The advantage of this base case is that it provides good control of P_H. However, from an energy utilisation point of view it is suboptimal to have the compressor work on and add compression work to an extra volume of gas, which subsequently is only used for condensation, and thus cooling, in order to balance the compression work. Energy would be saved by reducing the coolant flow rate by this extra amount. This increased energy efficiency is obtained by cooling in the top of the column, using a partial vapour condenser before energy is transferred to the heat pump, as shown by pinch analysis. An energy optimal flowsheet is shown in figure 6 where a partial internal reflux is generated by an internal cooler in the column top. The potential for energy saving with this design is revealed in table 2 with utility requirements for the base case and the optimal process design as well as for a non-energy integrated plant. The table shows that compared to the base case heat pump design, a 30% reduction in compressor energy consumption is obtained by applying the optimal design.

5.2.3 Stability and Operability

The change in stability properties of the optimally designed plant may be qualitatively deduced from the discussion of those of the base case design, since the energy flows Q_{cool} and W_{comp} are reduced with the optimal design. Thus the unstable real eigenvalue in focus will lie more to the left than for the base case. Thus the optimal design with the level loops closed also may be unstable but the unstable dynamics will be slower than that of the base case.
With the primary actuator control valve CV8 replaced by the valve CVi, one suggestion for the primary heat pump control configuration is: P_H is controlled by CV9, and P_L is controlled by CVi. With this configuration time delays and dynamics in the single loops are minimised. A somewhat lower control bandwidth compared to the base case must be expected due to slower interactions caused by the double heat exchange in the path from CVi to P_L. Thus the actuator levels shown in table 1-B are similar to those of the base case design with CV8 substituted by CVi as primary actuator and with the tertiary actuators interchanged.

Figure 6 illustrates how the column with the optimal cooler location balances between the temperature at the high pressure and a combination of the temperatures in the internal cooler (T_i) and at the low heat pump pressure. Switchability properties for this design will therefore be almost identical with those of the base case (figures 4–B and 4). The only deviation lies in the changes in P_L required to switch between operating points. These changes will be different since some of the driving force for the switches will stem from changes in T_i. From an operability point of view the main conclusion is that a major effect of the improved distillation column energy integration will be a reduced bandwidth for e.g. two product composition control. Thus in this case there exists a trade off between optimal energy utilisation and operability.

5.3 Optimisation Based Control Structure

Application of an optimisation method for control structure determination is exemplified here on the same column as studied by Hansen *et al.* (1998). Based on a benchmark problem formulation (Koggersbøl and Jørgensen, 1995) and assuming the above base level actuators, and control level to be available a product purity control structure is determined at a selected operating point. The method applied is an extension of that of Narraway *et al.* (1991). A number of stress levels are defined, where the purpose is to reject persistent oscillating disturbances in feed flow rate (fast) and feed concentration (slow). In this investigation is was decided to use the secondary actuator setpoints for the purity control. Thus the investigated problem was first to choose from the three actuators, $P_{H,s}$ $P_{L,s}$ and D_s, and 15 potential outputs a set of actuators and controlled variables to be included into the product purity control structure. For this selection perfect control is assumed. Second to determine which decentralised control configuration is the best for that particular set of measurements and actuators. The lower level control layer was already assumed implemented to stabilise the experimental distillation column. For the applied control structure selection methods a stable plant is required.
The result of the perfect control screening were five control structure (A-E) listed in table 3 and the existing structure F. These structures were evaluated using static and dynamic RGA, dynamic RDG, the presence of right half plane zeros and minimal condition number.

	output	actuator	cost
A	$T_{19},\ x_B,\ x_D$	$P_{H,s},\ P_{L,s},\ D_s$	0
B	$T_1,\ x_B,\ x_D$	$P_{H,s},\ P_{L,s},\ D_s$	0
C	$T_{19},\ P_{19},\ x_B$	$P_{H,s},\ P_{L,s},\ D_s$	$3\ 10^{-3}$
D	$T_1,\ P_{10},\ x_D$	$P_{H,s},\ P_{L,s},\ D_s$	$3\ 10^{-3}$
E	$T_1,\ P_{19},\ x_D$	$P_{H,s},\ P_{L,s},\ D_s$	$3\ 10^{-3}$
F	$x_B,\ P_{19},\ x_D$	$P_{H,s},\ P_{L,s},\ D_s$	0
G	$T_1,\ T_{19},\ P_{19}$	$P_{H,s},\ P_{L,s},\ D_s$	-

Table 3.: Selected structures by the perfect control selection method (A, B, C, D, and E). The existing structure (F). The result of realistic control screening method (G).

	Static RGA	Dyn. RGA	Dyn. RDG	RHP	CN
A	ic	ic	−	+	ic
B	ic	ic	−	+	ic
C	−	−	−	−	−
D	+	+	+	ic	ic
E	+	+	+	ic	ic
F	ic	ic	−	+	−

Table 4.: Summary of controllability measures for the structures selected by the perfect control screening. Symbols: − indicates structure is not favoured by the measure, + indicates structure is favoured, and 'ic' that the particular measure is inconclusive.

Table 4 summarises of the controllability measures for the 6 structures. The table suggests the overall qualitative ordering:

1	D,E
2	A,B
3	F
4	C

It is interesting to note that the RGA and RDG measures are the most conclusive indicators for the favoured structures, and that these measures all favour structures D and E. Thus structures D and E will most likely give the best control performance of the controlled variables.

5.3.1 Nonlinear Simulation

The structures A, D and F have been implemented in the nonlinear simulation program to verify the above result. Discrete time PI-controllers are used with a sampling time of 15 sec.

	x_D		x_B	
Struc.	min	max	min	max
A	-0.6	+0.6	-1.5	+0.7
D	-0.4	+0.5	-1.0	+0.6
F	-1.7	+1.0	-1.4	+1.3
G	-0.22	+0.17	-0.8	+0.5

Table 5. Minimum and maximum deviation on product purities, with stress level 2 oscillatory disturbances. A, D, and F are structures of the perfect control control screening, and D is of the realistic control screening

Table 5 shows the maximum and minimum time domain deviations from the steady state of the product purities when the column is subjected to the high stress level oscillatory disturbances. The simulations confirm the above ordering of the structures for best performance. The new structures A and D show improved performance compared to the existing structure F. Further it is seen that actually only structure D is able to meet the control objective of remaining within the allowable unity scaled deviations. Structure D may be viewed as an indirect control of the bottom product purity. Integral action could be achieved by a cascade using e.g. the setpoint of T_1 as manipulable variable. Whether this will improve performance compared to structure A or F in the step disturbance situation is to be investigated.

The second screening method was run with a 15 sec. delay on all measurements in order to attempt an automated tuning of the distributed PI-controllers. Just one feasible structure, G, was obtained. The automatic tuning did work, but it was far from optimal in the sense that is showed very oscillatory responses even to step disturbances. These responses were partially due to usage of Ziegler-Nichols tuning rules. For this reason the loops were detuned. The maximum and minimum deviation of the product purities when the system is subjected to the high amplitude disturbances are also given in table 5 and the product purity responses to a step feed flow rate disturbance are seen in figure 7. It is noteworthy that this structure gives far the best performance under the oscillatory disturbances and that it has an overshoot and settling time similar to those of structure D. That structure G was not among the selected structures of the perfect control screening indicates that this structure will not work with very tight control.

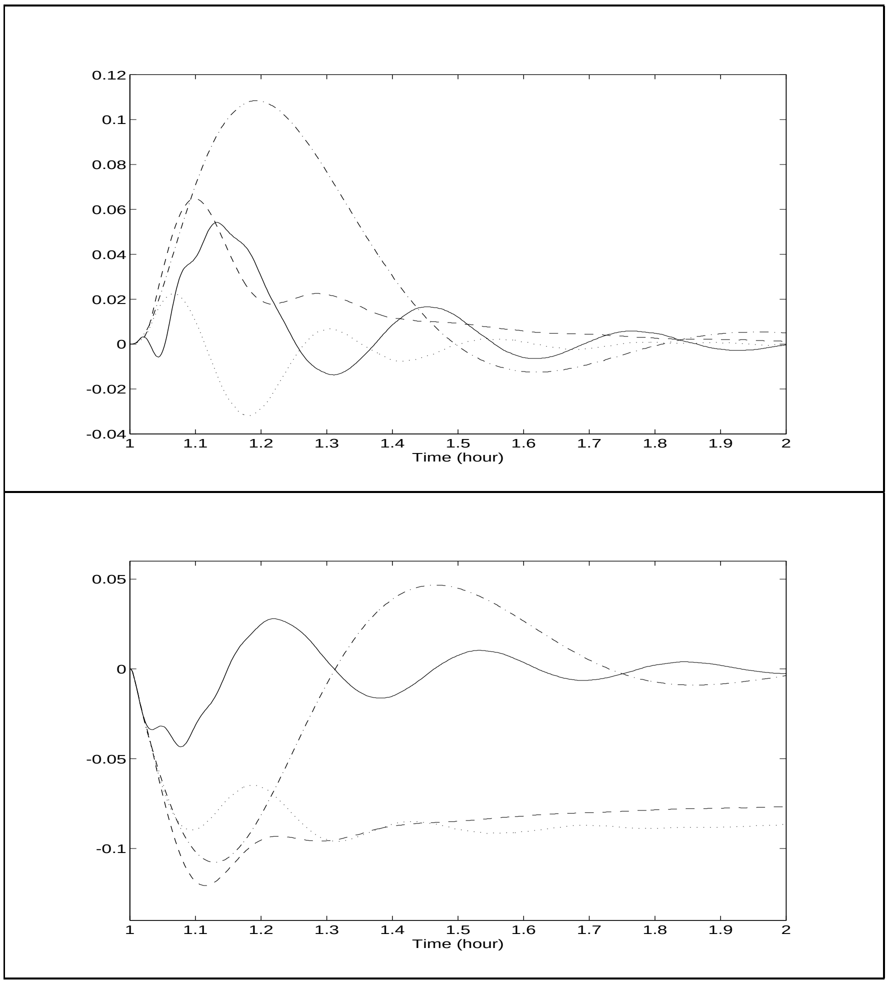

Figure 7. Responses to step disturbance in feed flow rate of 5%. Upper: x_D. Lower: x_B. solid: struc. A, dashed: struc. D, dotted: struc. G, dotdash: struc. F.

Discussion and Conclusions

The above case study indicates that a key aspect of control structuring is that of handling nonlinear phenomena during the different stages of a control structure design. It seems also clear that computer aided process engineering can provide significant assistance in integrating controllability considerations into plant design. An integrated computer aided system, ICAS (Gani *et al.*, 1997) has been developed which allows for application of integrated algorithms for controllability analysis, control structure development and process design. ICAS provides toolboxes for process design, process synthesis and process control which are integrated to each other and to a simulation engine. A process flowsheet can be simulated, designed and analysed in terms of controllability and verification of the control design can be carried out in an integrated manner in ICAS (ICAS manual 1999).
Integration of controllability considerations into plant design is limited by the complexity of the iterative plant design procedure. Some key issues are illustrated through the following five step

plant design procedure:

1. Conceptual plant design is subdivided into stages where design targets are set based upon thermodynamic constraints in each stage. Design targets are set according to product and waste quality requirements.

2. Design superstructure is proposed based upon mass and energy integration to suggest candidate plant configurations.

3. Nonlinear analysis is performed to evaluate the possibility of static multiplicities of candidate plant configurations, in order to decide upon possible switches between control configuration during operation.

4. Sketch design of proposed optimal alternatives for controllability evaluation near optimal operating conditions.

5. Selection of candidate for detailed engineering.

The steps selected in the above procedure are illustrative of some of the many decisions which have to be made during process design. Most often each step is carried out several times during a design. The above simplified procedure may be viewed as representing the evolution from more traditional design procedures, where only little optimisation and no process integration might have been achieved during the often very iterative design to more research level design procedures where a design assistant tool may support the designer in making the many design decisions in a consistent manner and where optimisations also guide the designer to reduce the need for iterations.

To include controllability considerations into process design require the following types of controllability evaluation tools:

1. Nonlinear analysis to reveal regions with qualitatively different behaviours.

2. Qualitative controllability preferably both at at a total plant level and at each stage of process design.

3. Quantitative controllability evaluation at each stage where models are available.

4. Plant wide controllability evaluation.

At the two latter levels 3 and 4 the tools reviewed above may be used, whereas tools for qualitative controllability evaluation could be e.g. Hopkins *et al.* (1998). Linear model based controllability analysis methods may be incorporated into optimisation based process design steps. To include linear model based analysis tools into qualitative step might be achieved through combining thermodynamically based qualitative models with qualitative information from linear qualitative models. However when linear analysis is used, this analysis should be preceded by a nonlinear analysis to reveal where within the operating window there are possibilities for multiple solution. Within each of these regions different control structures may be needed. Transitions between such operating regions require special attention.

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