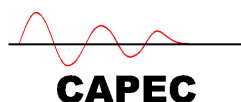


Computer Aided Process Engineering for Operational Analysis and Control Design

Torben Ravn Andersen and Mario Richard Eden



Computer Aided Process Engineering Center
Department of Chemical Engineering
Technical University of Denmark
DK 2800 Lyngby, Denmark

Abstract

The emphasis of this paper is upon computer aided analysis of process dynamics, operability, process control structuring and design, whereas Eden and Andersen (2000) covers property prediction, computer aided molecular design, process synthesis/design and process integration.

1. INTRODUCTION

By definition Computer Aided Process Engineering (CAPE) refers to computer aided methods, algorithms, and techniques related to process and product engineering. This contribution aims at describing some achievements of the research conducted in this area by the Computer Aided Process Engineering Center (CAPEC) at the Department of Chemical Engineering, Technical University of Denmark. The research in CAPEC is divided into seven research programs, where research program A deals with modelling of properties (phenomena), while program B deals with modelling and identification of processes, operations and phenomena not covered in research program A. Based on the models from research programs A and B, algorithms, methods and computer-aided tools for synthesis, design and analysis of processes, products and operations (research program C) are developed. Simultaneously, the developed models provide the basis for research in process operation and control (research program D). The results from research programs A-D provide the foundation for research programs E (process and tools integration) and G (safety and hazards). Since all research programs need numerical tools and databases, research program F supplies this need to all levels. Based on the above outlined description, the research objectives of CAPEC can be summarized as:

Develop computer-aided systems for process simulation, process/product design and analysis, control/operation principally suitable for chemical, petrochemical, pharmaceutical and biochemical industries. The computer-aided systems are to be developed based on fundamental modelling studies that incorporate correlation and estimation of thermo-physical and phase equilibrium properties as well as modelling the underlying principles/behaviour of the process.

The emphasis of this paper will be upon computer aided analysis of process dynamics, operability, process control structuring and design, whereas Eden and Andersen (2000) covers property prediction, computer aided molecular design, process synthesis/design and process integration.

2. Modelling and Identification

The idea of development or identification of mathematical models for process units or process operations is to compress the significant information in a suitable reusable form. The model may be used to improve the understanding of the process/operation through experiments and/or simulation. The models can be applied to design, control and optimisation. Thus the modelling purpose may be a combination of desires to save time and money, to decrease harmful emissions to the environment and to ensure a consistent and reliable production.

Modelling and identification may be divided into three related types of projects:

1. Phenomena based modelling
2. Data based modelling
3. Development of modelling techniques

The phenomena based modelling starts with first principles models and adds empirical correlations where necessary, whereas data based modelling uses experimental data combined with process knowledge. Thus the main difference between the two approaches is the starting point, namely process knowledge versus data. The phenomena based models generally provide predictive models whereas data based models traditionally are correlations, with no predictive capabilities.

In CAPEC modelling, identification and simulation are carried out within the following fields:

1. Separation processes
2. Reactors, bioreactors and reaction pathways
3. Biological waste water treatment processes
4. Heat-integration systems
5. Batch and Fed-batch operation (phenomena based)
6. Data based modelling

In the following three small examples will be given within the fields of separation process modelling and reactor modelling.

2.1. Example 1: Heterogeneous Azeotropic Distillation Sequences

The static operation of heterogeneous azeotropic distillation sequences have been investigated. This subject area is of significant interest since heterogeneous distillation usually takes place at lower temperatures than homogeneous distillation, and thus is more interesting seen from an energy efficiency point of view.

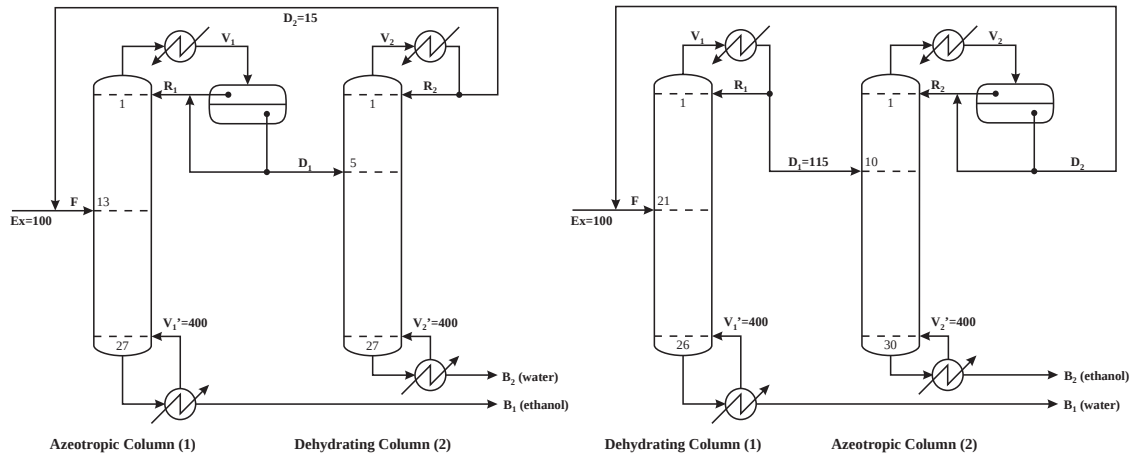


Figure 1: Direct sequence

and Indirect sequence

The simulation was carried out in ICAS and it was found that multiple steady states exist within a broad range of the feasible operating region, Esbjerg et al. (1998). Figure 1 outlines 2 different distillation sequences suitable for ethanol dehydration (with addition of cyclohexane as an entrainer).

2.2. Example 2: Fixed Bed Reactor with Mass and/or Energy Recycle

In this example the main points of investigation have been analysis of a simple first order reaction in a fixed bed reactor with either reactant recycle and/or energy recycle. The focus was on the complex dynamic behaviour that can arise in these types of systems and on comparison of different ways to achieve recycle. The reacting systems were modelled by coupled partial and ordinary differential equations (and in some cases also algebraic equations). The results confirmed that it was possible to find multiple periodic solutions as well as multiple steady states in a fixed bed reactor when recycle was applied, Recke (1998).

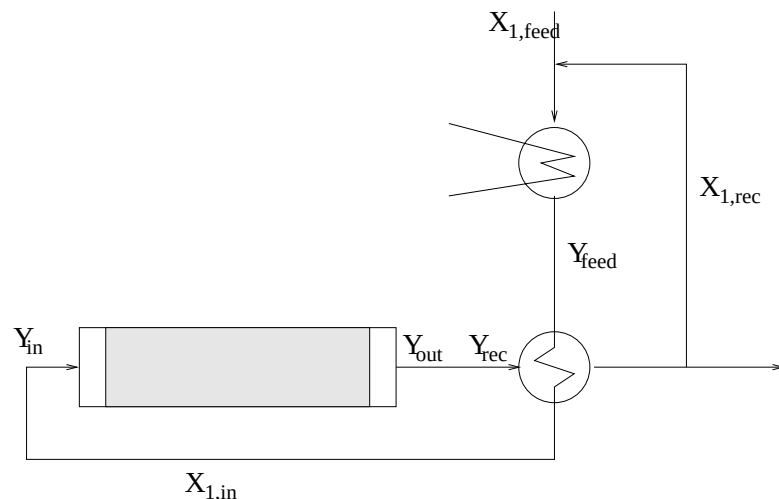


Figure 2: Fixed bed reactor with mass and/or energy recycle

The above project was initiated after obtaining experimental evidence of the existence of damped periodic solutions in an experimental fixed bed reactor with reactant recycle. With the above investigation it is now of interest to further investigate experimentally the presence of the predicted bifurcation scenario.

2.3. Example 3: Catalytic Fixed Bed Reactor with Energy Recycle

A reactor system consisting of a catalytic fixed bed and a feed-effluent heat exchanger has been investigated. The reactor system produces ammonia. A bifurcation analysis of the process model has shown that the operating region is divided into regions with different (dynamic) behaviour. See further description of the dynamic behavior below.

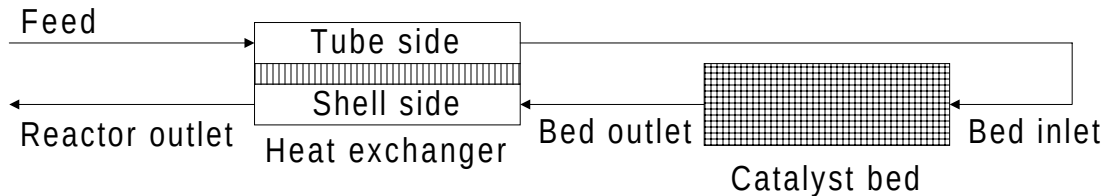


Figure 3: Sketch of the ammonia reactor system.

3. Operability Analysis

The integration of processes can give new and significantly more efficient plants seen from an energy as well as an economical point of view, BUT process integration typically also implies advanced or complex behaviour statically as well as dynamically. It often gives rise to:

- **Multiple static operation modes**
 - *Output multiplicity*
The existence of multiple static solutions in terms of the outputs (dependent variables) when the inputs (independent variables) are specified.
 - *Input multiplicity*
The existence of multiple static solutions in terms of the inputs when the outputs are specified.
 - *(Internal) State multiplicity*
The existence of multiple static solutions in terms of the internal states when the inputs and outputs are specified.
- **Multiple periodic solutions**
 - *Stable periodic solutions*
Where the operation is sustained oscillations.
 - *Unstable periodic solutions*
Where the operation has oscillatory transients and eventually may end up on a new stable solution or possible blow-up.

Examples of applications with multiple steady states and multiple periodic solutions are many. It includes for example fixed bed reactors with recycle, binary distillation, azeotropic distillation. For some of the mentioned applications see e.g. Recke (1998), Jacobsen and Skogestad (1991), Koggersbøl et al. (1996), Petlyuk and Avet'yan (1971), Bekiaris et al. (1993).

In the following operability aspects for the examples of heterogeneous azeotropic distillation sequences and ammonia reactor system are outlined.

3.1. Example 1 (cont'd): Heterog. Azeotropic Distillation Sequences

The figure below shows that multiple steady states exist in a broad range of the operating region for the direct sequence. Also it is seen that input multiplicity exist for both sequences.

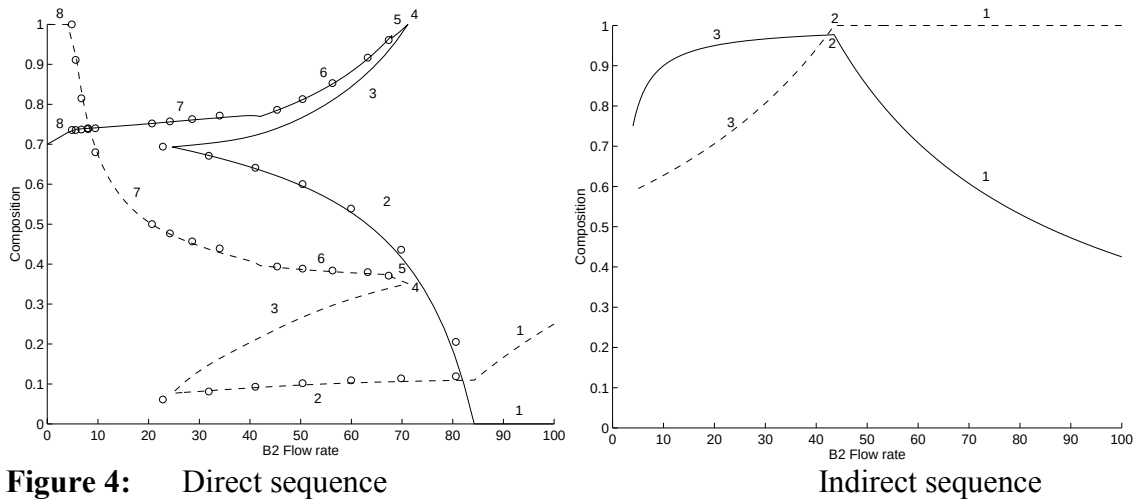


Figure 4: Direct sequence

Indirect sequence

3.2. Continuation

In CAPEC a continuation program is being integrated with an in-house developed dynamic process simulator within the Integrated Computer Aided System (ICAS). The continuation program has been developed by the CONT Organization at the Department of Chemical Engineering, Prague Institute of Chemical Technology. For details on the program and methods used within this framework see Kubicek (1976), Kubicek and Marek (1983), Schreiber and Marek (1991 & 1995).

A continuation program searches a dynamic model for solutions as one or more of the model parameters (termed bifurcation parameters) are varied. A bifurcation point is a point within the state and parameter space where the number and/or the stability of solutions changes as the bifurcation parameter is varied in the neighborhood of the point.

3.3. Example 3 (cont'd): Catal. Fixed Bed Reactor w. Energy Recycle

In the ammonia reactor example described above, two fold bifurcation points has been found to limit a region with three steady states, one ignited, one extinct and one constantly unstable in-between. The ignited steady state has been found contain a stable part and an unstable part divided by a Hopf bifurcation. The Hopf bifurcation has been found to be sub-critical and an unstable periodic solution emerged, coexisting with the stable ignited steady state. This unstable periodic solution was found to undergo a cyclic fold bifurcation and turned unstable. Consequently, a region bounded by the Hopf bifurcation on the ignited steady state and the cyclic fold bifurcation had at least two stable steady states, one unstable steady state, one unstable periodic solution and one stable periodic solution. This region included the operating point with the highest production rate. This analysis pointed to consequences of operation of an uncontrolled reactor system since a disturbance may move the operating point from the desired stable steady state to the stable periodic solution.

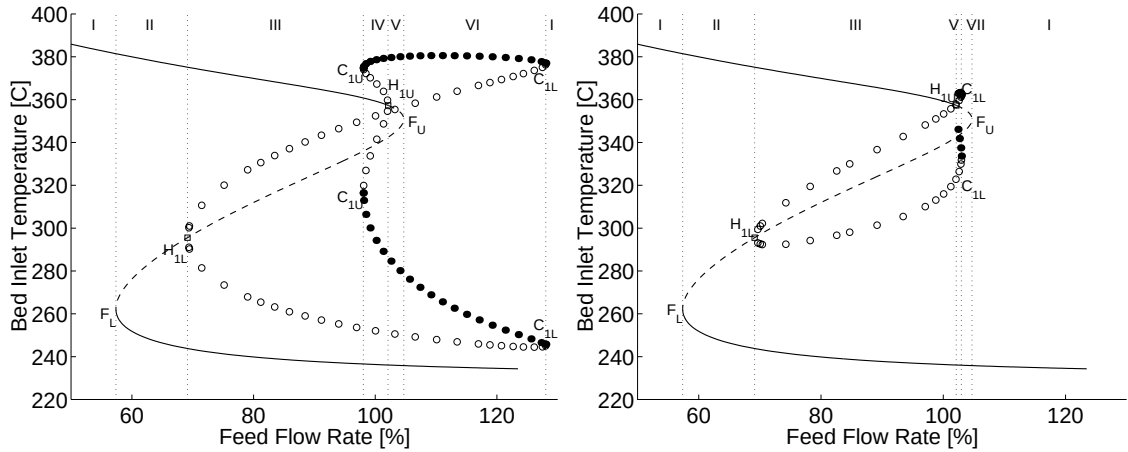


Figure 5: Bifurcation diagrams for catalytic fixed bed reactor with energy recycle. Right figure: Open loop. Left figure: Closed loop (bifurcation control – see below!).

4. Process Control and Operations

From the valuable knowledge of process dynamics obtained by using the continuation analysis tool, control structure selection and design can be pursued. In CAPEC research in process control and operations are carried out within the following fields:

1. Batch process identification and control
2. Non-linear dynamics and control
3. Plant wide control
4. Software implementation

In the following section examples from the non-linear control area as well as the plant wide control field (control structure selection) are given.

4.1. Example 3 (cont'd): Catal. Fixed Bed Reactor w. Energy Recycle

Based upon the fundamental understanding of bifurcations in recycle fixed bed reactors developed, a bifurcation based controller has been developed. The aim is to prevent the onset of sudden large amplitude temperature and concentration oscillations, which have been reported in literature to have occurred in practice (Andersen, 1999).

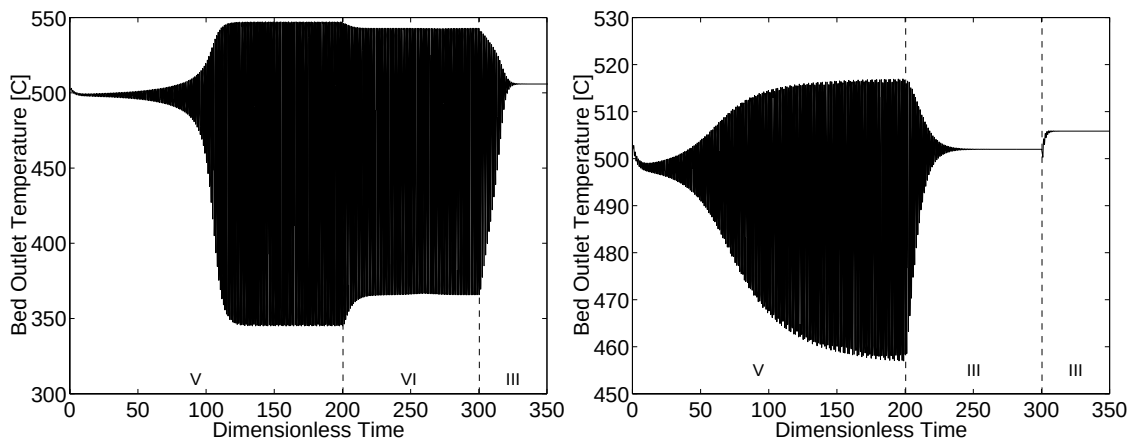


Figure 5: Simulation with feed flow rate step disturbance at $t = 0$. Right figure: Open loop (with hysteresis). Left figure: Closed loop (without hysteresis).

The bifurcation-based controller has been implemented on the simulated ammonia reactor. Simulation with this control can be seen in figure 6. The regulator achieved the expected goal of preventing the onset of large-scale oscillations (Recke *et al.*, 2000).

4.2. Control Structure Selection

In process design it is ideally necessary to simultaneously consider the production quality, productivity, process safety, energy and resource saving. Thus the concept of generation of energy and resource saving chemical engineering processes presupposes integrated synthesis of equipment design and control structure design. The traditional approach is to design a plant for some desired nominal operating regimes, which can be found during experimental research on a pilot plant. However this approach is troublesome due to the large number of uncertain disturbances which are not taken into account and which appear in practice. Therefore sequential (multistage) design of the process units and control structure may lead to decreasing economic benefit as a result of controllability problems for the process. That is the process may not be able to operate at the optimal productivity / production rate – or worse - it may not be able to operate at all. Integrating process design with control structure design can give a plant that is more easy to operate/control and hence an overall more optimal plant, see Jørgensen *et al.* (1999).

A control structure selection strategy based on a linear state space model is used to determine economically optimal pairings of controlled and manipulated variables. To provide optimal selection of the control structure, an MILP formulation based on the smallest total relative interaction has been used. After the relaxation of the binary restrictions this optimization problem can be treated as an LP problem with the solution identical to the original MILP problem. This tool has been implemented. The aspect of controller tuning can also be addressed.

4.3. Example 4: Energy Integrated Distillation Column

Application of controllability analysis for control structuring have been exemplified on a number of examples. One study object is an 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.

4.3.1. Plant Description

Figure 6 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 (Throttle

valve) throttles high pressure liquid refrigerant to the heat pump low pressure (P_L) suitable for evaporation in the condenser. The control valve (CV_9) 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 CV_8 . Through a storage tank (Rec) and a heat exchanger the refrigerant cycle is closed at the expansion valve.

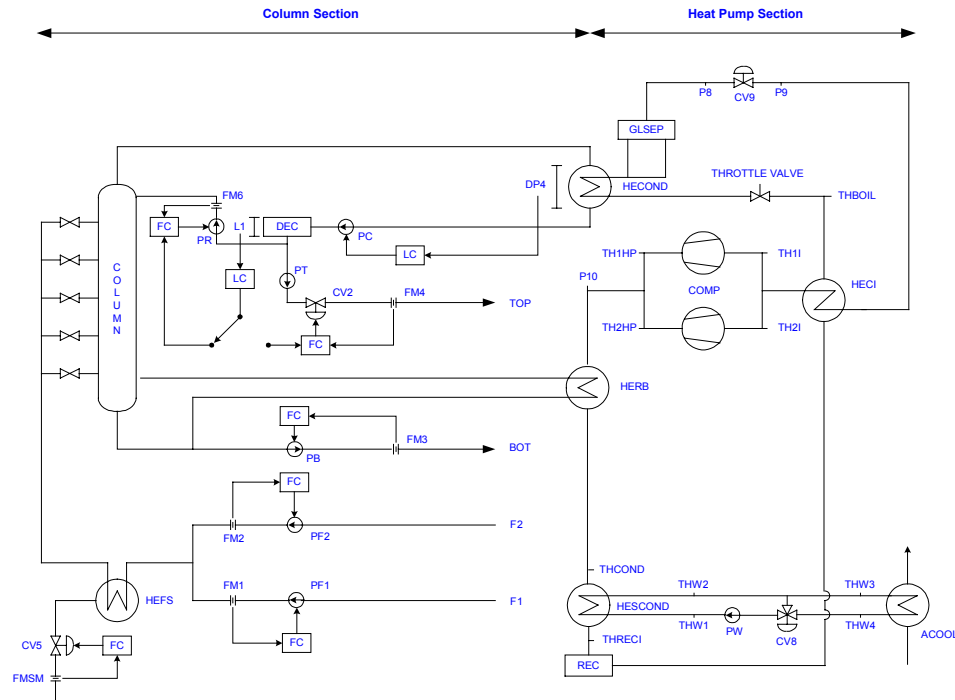


Figure 6: Schematic flowsheet of the heat integrated distillation plant at department of chemical engineering, DTU.

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.

4.3.2. 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.

4.3.3. 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 CV_8 or the throttle valve CV_9 . Since CV_8 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 CV_8 to stabilise the system. However, the gains from CV_8 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 CV_9 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 CV_8 . 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 CV_9 . Since the plant is already stabilised by the CV_8 - P_H loop CV_9 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 CV_8 - P_H loop.

By controlling the conditions on the refrigerant side of the two contact points using the CV_8 and CV_9 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.

4.3.4. 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.

Column	Primary	Secondary	Tertiary
Top	D L		CV_9 CV_8
Heat pump	$P(P_H + P_L)$ $V(P_H - P_L)$	P_L P_H	
Bottom	B		

Table 1: Actuator levels for 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:

$$\begin{aligned}
 V \cdot \rho_l \cdot \Delta H^{vap} &= AU_B \cdot (T_{refrig}^{sat}(P_H) - T_B) = AU_C \cdot (T_C - T_{refrig}^{sat}(P_L)) \\
 P_C &= g_1(T_C) \\
 P_B &= g_2(T_B) = P_C + \Delta P_{trays}(V)
 \end{aligned}$$

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. $T_{refrig}^{sat}(P_j)$ 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. The equations can be interpreted as shown in figure 7-A.

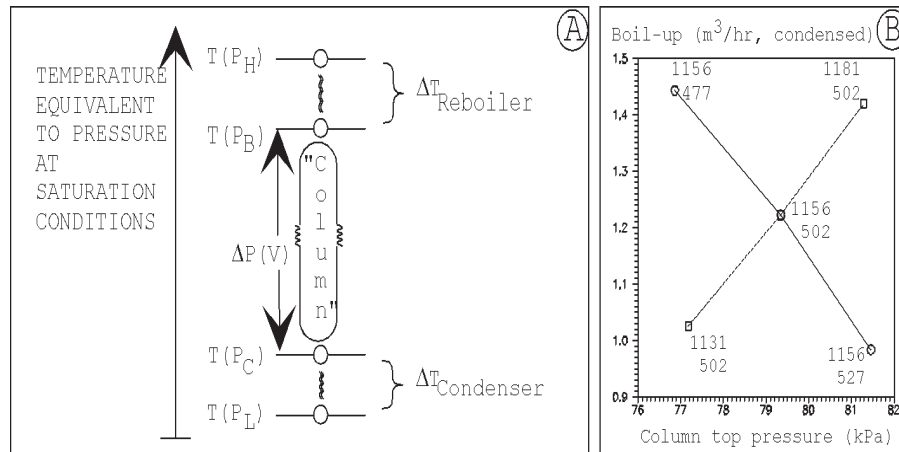


Figure 7: A: The distillation column modelled as an elastic body "balanced" (in "springs") between the temperatures corresponding to the actuators P_H and P_L . The "springs" inside the column symbolise that the pressure drop from bottom to top can vary as a function of vapour flowrate. B: Steps ± 25 kPa in high and low heat pump pressures (simulation) (on plot: P_H over P_L in kPa)

When P_H and P_L are brought to specified values by the CV_8 - and CV_9 -controllers the system settles at values for boil-up and column pressure(s) such that the column "balances" between the two heat pump pressures.

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 can be shown by pinch analysis.

4.4. 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 it 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 structures (A-E) listed in table 2 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.

Structure	Output	Actuator	Cost	Static RGA	Dyn. RGA	Dyn. RDG	RHP	CN
A	T_{19}, x_B, x_D	$P_{H,s}, P_{L,s}, D_s$	0	ic	ic	-	+	ic
B	T_1, x_B, x_D	$P_{H,s}, P_{L,s}, D_s$	0	ic	ic	-	+	ic
C	T_{19}, P_{19}, x_B	$P_{H,s}, P_{L,s}, D_s$	$3 \cdot 10^{-3}$	-	-	-	-	-
D	T_1, P_{10}, x_D	$P_{H,s}, P_{L,s}, D_s$	$3 \cdot 10^{-3}$	+	+	+	ic	ic
E	T_1, P_{19}, x_D	$P_{H,s}, P_{L,s}, D_s$	$3 \cdot 10^{-3}$	+	+	+	ic	ic
F	x_B, P_{19}, x_D	$P_{H,s}, P_{L,s}, D_s$	0	ic	ic	-	+	-
G	T_1, T_{19}, P_{19}	$P_{H,s}, P_{L,s}, D_s$	-					

Table 2: 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). 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.

The results seen in table 2 summarises the controllability measures for the 6 structures, and suggests the overall qualitative ordering:

best	1	D, E
	2	A, B
	3	F
worst	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.

4.4.1. Non Linear Simulation

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

Structure	x_D		x_B	
	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 3: 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 G is of the realistic control screening.

Table 3 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_I 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 it 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 3 and the product purity responses to a step feed flow rate disturbance are seen in figure 8. 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.

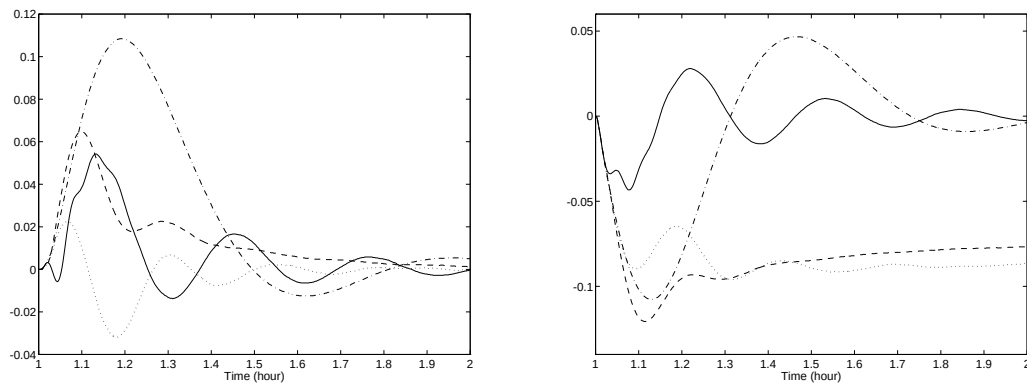


Figure 8: Responses to step disturbance in feed flow rate of 5%. Left: x_D (deviation) Right: x_B (deviation). solid: structure A, dashed: structure D, dotted: structure G, dotdash: structure F.

5. Conclusions

An overview of the research efforts of the Computer Aided Process Engineering Center (CAPEC) has been presented. The emphasis of this paper has been on computer aided analysis of process dynamics, operability, process control structuring and design. Several examples have been given in order to illustrate some of the developed methodologies, tools and algorithms.

6. Acknowledgement

Although the results presented in this paper are the achievements of all CAPEC coworkers, a few persons need to be highlighted. The directors of CAPEC, Professor Sten Bay Jørgensen and Dr. Rafiqul Gani for their research vision and leadership. Bodil Recke for assisting with the description of the reactor examples.

References

1. B.R. Andersen, "Non-Linear Dynamics of Catalytic Ammonia Reactors", Ph.D. thesis, 1999.
2. N. Bekiaris, G.A. Meski, C.M. Radu, M. Morari, "Multiple Steady States in Homogeneous Azeotropic Distillation", I&EC Research, 32, pp 2023-2038, 1993.
3. I-Lung Chien, and P.S. Fruehauf (1990): "Consider IMC Tuning to Improve Controller Performance", Chem.Eng.Progress, vol. 86(10), pp.33-41.
4. M.R. Eden and T.R. Andersen, "Computer Aided Process Engineering For Process Integration And Design", submitted for SIMS 2000, Technical University of Denmark, 2000.

5. K. Esbjerg, T.R. Andersen, S.B. Jørgensen, D. Müller and W. Marquardt, "Multiple Steady States in Heterogeneous Azeotropic Distillation Sequences", I&EC Research, 37, pp 4434-4452, 1998.
6. J.E. Hansen, J. Heath, J. Perkins and S.B. Jørgensen (1998), "Control Structure Selection for Energy Integrated Distillation Column", J.Pro. Cont., 8, pp. 185-195.
7. E.W. Jacobsen and S. Skogestad, "Multiple Steady States in Ideal Two-Product Distillation", AIChE Journal, vol 37(4), pp.499-511, 1991.
8. S.B. Jørgensen, R. Gani and T.R. Andersen, "Towards Integration of Controllability into Plant Design", MED'99, Haifa, Israel, June 1999.
9. A. Koggersbøl and S.B. Jørgensen (1995a), "Distillation Column Control Benchmarks: Four Typical Problems", DYCORD+95, Helsingør, Denmark, June 5-7th, pp.323-326, 1995.
10. A. Koggersbøl and S.B. Jørgensen (1995b), "Validation of a Distillation Column Benchmark", Proceedings of IFAC Symposium DYCORD+95, Helsingør, Denmark, pp.509—514, 1995.
11. A. Koggersbøl, B.R. Andersen, J.S. Nielsen and S.B. Jørgensen (1996a), "Control Configurations for an Energy Integrated Distillation Column", Comp. Chem. Eng., 20S, pp. S853-S858.
12. A. Koggersbøl, T.R. Andersen, J.B. Jørgensen, S.B. Jørgensen (1996b), "An Output Multiplicity in Binary Distillation: Experimental Verification", Computers & Chem Eng, 20 (supplement), pp.S835-S840, 1996.
13. M. Kubicek, Algorithm 502. Dependence of solutions of nonlinear systems on a parameter, ACM Trans. Math. Software 2, 98, 1976.
14. M. Kubicek and M. Marek, Computational methods in bifurcation theory and dissipative structures, Springer-Verlag, New York, 1983.
15. L.T. Narraway, J.D. Perkins and G.W. Barton (1991), "Interaction between process design and process control: Economic analysis of process dynamics", Journal Process Control, 1, pp. 243-250.
16. F.B. Petlyuk and V.S. Avet'yan (1971), "Investigation of the Rectification of Three-component mixtures with Infinite Reflux", Theor. Found. Chem. Eng., 5, pp. 499-507, 1971
17. B. Recke, "Non-Linear Dynamics and Control of Chemical Processes", Ph.D. thesis, 1998.
18. B. Recke, B.R. Andersen and S.B. Jørgensen, "Bifurcation Control of Sample Chemical Reaction Systems", ADCHEM 2000, June, Italy, 2000.
19. I. Schreiber and M. Marek, Chaotic behaviour of deterministic dissipative systems, Cambridge Univ. Press, Cambridge, 1991 & 1995.