INTEGRATION OF DESIGN AND CONTROL FOR ENERGY INTEGRATED DISTILLATION

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ABSTRACT

A systematic computer aided analysis of the process model is used as a pre-solution step for integration of design and control problem. In this paper, a static energy integrated distillation plant model is first presented, then the analysis is presented for a single distillation column, subsequently the column analysis is extended with the analysis for the heat pump. The analysis relates to the process and control design. The sensitivities of the plant are revealed in the analysis and through simulation. The insights developed hereby lead to a modification of the actuator set for the integrated plant. The final result holds the promise that an integrated design and control problem can be defined correctly and consistently with respect to the process behaviours and the selected process models. The key contribution of this work is the qualitative analysis and insights gained from the model analysis with respect to actuator selection and to control structure selection.

INTRODUCTION

Chemical process plant design and operation based on combining mathematical models with computer science have the potential to significantly increase the efficiency of manufacturing systems by integrating the design with the planning of operation. At the level of chemical plant units, integrating systems means coupling several physical processes either within the same piece of equipment or between units, which perform different physical/chemical tasks. Full economic potential of such novel integrated processes can only be exploited if they can be operated efficiently.

Simultaneous design of process integration on distillation columns and analysis of the control problem has significant practical implications. Furthermore distillation plants exhibit some interesting and operationally highly relevant multivariable dynamics. Understanding such characteristics is essential for being able to handle multivariable control issues, which again is a premise for being able to optimise the operation of chemical plants.

Therefore it is essential to investigate operability of such integrated processes at the design stage in order to enable suitable design modifications before it is too late. A systematic computer aided pre-solution analysis of the process model analysis method for integration of design and control presented by Russel et al. (2002)[1] is investigated on a process integrated plant. The process model equations are classified in terms of balance equations, constitutive and conditional equations. Analysis of the phenomena models, which represent the constitutive equations identify relationships between important process and design variables. These relationships help to understand, define and address issues related to integration of design and control decisions. The analysis enables identification of a set of process (control) variables and a set of design (actuator) variables that may be employed with different objectives in a process design context versus in a control actuator design context. Consequently these sets play an important role in formulating an integrated design and control problem (Russel et al (2002) [1] and Tyreus(1999)[2]). In partially related works, Agrawal and Fidkowski (1999)[3] have proposed new thermally coupled schemes from a steady state analysis only, while Hernández and Jiménez (1999)[4] have analysed the controllability of thermally coupled distillation columns.

This paper investigates design and control of an integrated distillation column through an analysis of the model equations and validates the results with simulation analysis of the energy integrated distillation column. The energy integrated example used in this work constitutes the tightest possible heat pump integrated distillation example. In industrial practice often several columns may be integrated with one heat pump; the present system presents a tighter coupling than those.

First the energy integrated distillation plant model is presented, then the analysis is presented for a single distillation column, subsequently the column analysis is extended with the analysis for the heat pump. The analysis directly leads to definition of an operating window for the distillation column, wherein design optimisation may be carried out, and wherein the heat pump should operate the column, using a set of actuators designed as a part of the heat pump. The sensitivities of the plant are revealed in the analysis through simulation of the plant. The insights developed hereby lead to a modification of the actuator set for the integrated plant.

MODEL FORMULATION

The starting point for the model analysis is to use an appropriate process model. The heat integrated distillation pilot plant, shown in Figure 1[5], contains two main parts, namely a distillation column section and a heat pump section. The heat pump section is physically connected to the distillation column through the condenser at the top and the reboiler at the bottom of the column. A more detailed description is available in Eden et al. 2000[6]. The process model equations of these two sections are discussed individually.

Process Model Equations for Distillation Column

The steady state process model equations for the distillation column are given below: Component balance equations (on tray j, j=1,NP; where NP is the total number of trays):

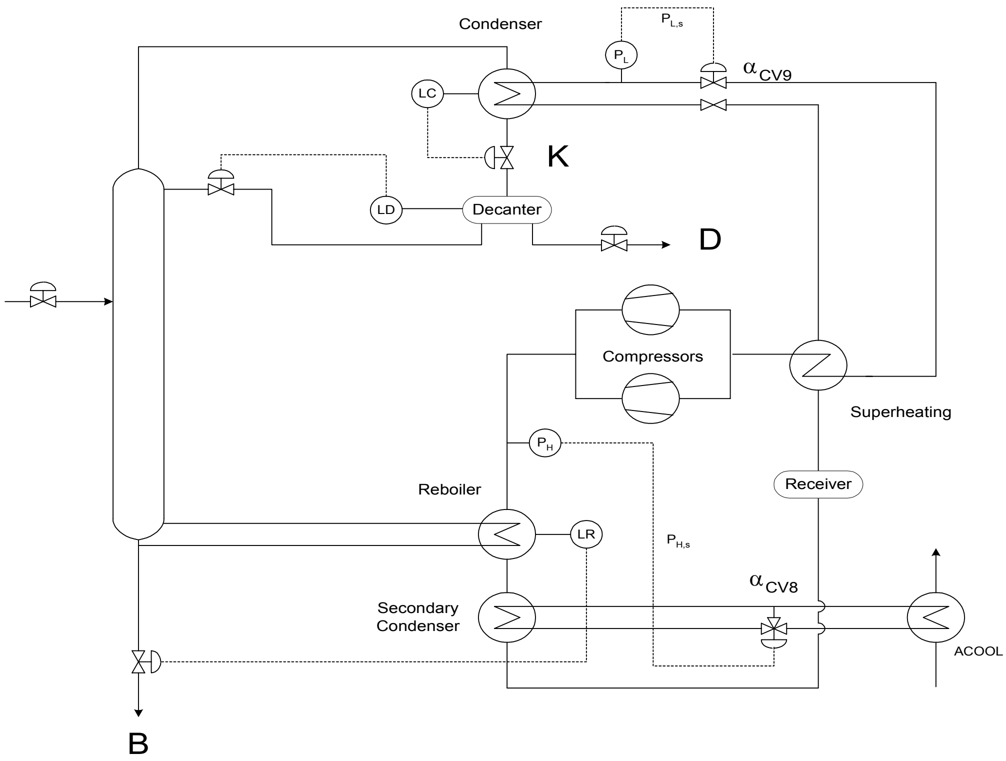

Figure 1: Flowsheet for the heat integrated pilot plant

$$0 = F_j Z_{i,j} + L_{j-1} x_{i,j-1} + V_{j+1} y_{i,j+1} - L_j x_{i,j} - V_j y_{i,j} \quad \text{i=1,NC; j=1,NP} \qquad (1)$$

Where, for total condenser,

$$0 = V_1 - (L_0 + D) \qquad (2)$$

Where, for reboiler,

$$0 = L_{NP} - (V_B + B) \qquad (3)$$

Energy balance:

$$0 = F_j H_j^F + V_{j+1} H_{j+1}^V + L_{j-1} H_{j-1}^L - V_j H_j^V - L_j H_j^L \quad \text{j=1,NP} \qquad (4)$$

$$0 = V_1 H_1^V - (L_0 + D) H_D^L - Q_C \qquad \text{(Total condenser)} \qquad (5)$$

$$0 = (L_{NP} - B)H_{NP}^{L} - VH_{B}^{V} + Q_B \qquad \text{(Reboiler)} \qquad (6)$$

Constitutive equations:

$$H_j^L = f_1(T, x_{i,j}) \qquad \text{j=1,NP} \qquad (7a)$$
$$H_D^L = f_2(T, y_1) \qquad \text{j=1,NP} \qquad (7b)$$
$$H_j^V = f_3(T, y_{i,j}) \qquad (8a)$$
$$H_B^V = f_4(T, x_{NP}) \qquad (8b)$$
$$y_{i,j} = K_{i,j} x_{i,j} \qquad \text{i=1,NC; j=1,NP} \qquad (9a)$$
$$K_{i,j} = f_5(T, P, x_{i,j}, y_{i,j}) \qquad (9b)$$

Conditional equations (j=1,NP)

$$0 = \sum_{i=1}^{NC} x_{i,j} - 1 \quad \text{i=1,NC} \qquad (10)$$
$$0 = \sum_{i=1}^{NC} y_{i,j} - 1 \quad \text{i=1,NC} \qquad (11)$$
$$P_j - P_{j-1} = \Delta P \qquad (12a)$$

In the above models, Eq.(1)-Eq.(3) represent the mass balance on each tray for NC components in the column, Eq.(4)-Eq.(6) represent the energy balance on each tray, where Q_c and Q_B are the heat removed in the condenser and heat added in the reboiler, respectively. Eq(7a)-Eq(9b) represent the phenomenon models for enthalphy, gas-liquid equilibrium condition and the so called equilibrium constant. Eq.(10) and Eq.(11) represent the conditional equations. Eq.(12a) represents the pressure drop on each tray. Subscripts i and j represent component and tray respectively.

The variables (P, T, $x_{i,j}$) are assigned as the process variables that also are the intensive variables. The possible optimization variables are L_0, V_B, P_C, Q_C, Q_B, D, B, while the constitutive variables are $K_{i,j}$ and $H_{i,j}$.

Process Model for the Heat Pump

Figure 2[7] shows a schematic of the heat pump side. The heat pump consists of four heat exchangers, two compressors, one expansion valve, a large tank, α_{CV8} and α_{CV9} are the control valves. While the refrigerant circulates within the heat pump it changes phase, and through absorbing heat of vaporization at low pressure and releasing it again at high pressure it carries heat from the column condenser to the column reboiler. A static model of the heat pump is presented for the purpose of this analysis.

From the energy balance, the heat duty for the heat condenser:

$$Q_C = f_6(L_{EXP}^H, \Delta H_{vap}^H(T,P), Q) \tag{12b}$$

where: L_{EXP}^H is the flow of refrigerant through the expansion valve;

$\Delta H_{vap}^H(T,P)$ is the vaporization heat of refrigerant.

From the energy balance over the expansion valve, the flow rate:

$$L_{EXP} = f_7(c_p, \Delta H_{vap}^H, P_H, P_L) \tag{13}$$

The heat duty for the reboiler:

$$Q_B = f_8(L_{EXP}^H, \Delta H_{vap}^H(T,P), Q) \tag{14}$$

The heat duty for the vapour superheater before compressor:

$$Q = f_9(T_3, T_6) \tag{15}$$

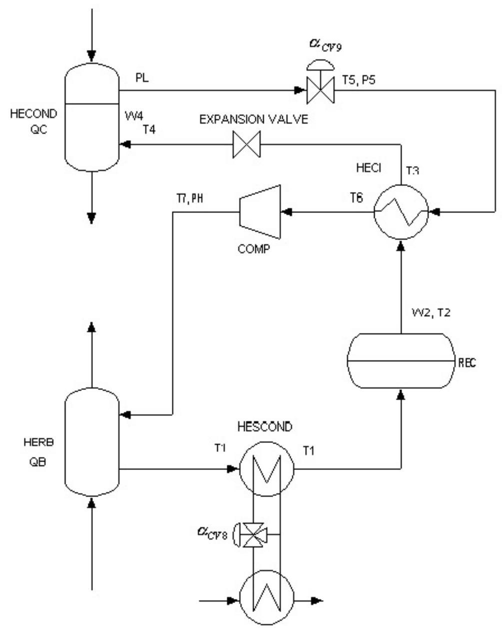

Figure 2: Flowsheet of the heat pump

From Eq.(12b)-Eq.(15) Q_B and Q_C on the heat pump side are functions of geometric size of equipment and the properties of the heat pump fluid. For the selected equipment, they are only a function of the property of the heat pump fluid. The design and control of the integrated distillation column rely on the properties of the heat pump fluid for the selected equipment. Hence the design and control issue of the integrated distillation column depends upon the properties of the refrigerant on the heat pump side.

PROCESS MODEL ANALYSIS

Model analysis first determines the available degrees of freedom for design and for control. This paper will consider static aspects, thus the number of degrees of freedom is the same for design and for control. The analysis subsequently identifies the important constitutive equations, their dependent process variables and the corresponding derivative information. The analysis of the balance equations and the constitutive equations generate information related to process sensitivity, process feasibility, design constraints and provide solutions for design and control sub-problems. The model analysis steps include problem definition, model construction, model solution and model validation through combining analysis and rigorous simulation. First the analysis is carried out for the distillation column; thereafter the heatpump is introduced into the analysis.

Process Model Analysis for the Distillation Column

In the above distillation column model, there are 4NC*NP+8NP+11 variables ($x_{i,j}$, $y_{i,j}$, $z_{i,j}$, $K_{i,j}$, F_j, $H^F{}_j$, L_j, V_j, $H^V{}_j$, $H^L{}_j$, T_j, P_j, L_0, V_B, P_C, ΔP, $H^L{}_D$, $H^V{}_B$, Q_C, Q_B, D, B, NP), 3NC*NP+6NP+6 independent equations(Eq.1-Eq.12a). So degrees of freedom are NP*NC+2NP+5. For one feed, the number of variables are 3NP*NC+NC+6NP+13+N_F, where N_F is the location of feed plate. For this case, the degrees of freedom become NC+7+N_F. If the variables related to the feed, the number of trays and the pressure drop are considered known ($z_{i,j}$, F, H_F, NP, ΔP, N_F). There are three variables that need to be specified from L_0, V_B, P_C, Q_C, Q_B, D, B in order to solve the 3NC*NP+6NP+6 independent equations. This means that in this distillation column three variables (design variables) are related to three process variables (controlled variables). With these three variables specified, all other variables in the system can be calculated.

From Eq.(9b) one can see that the equilibrium constants $K_{i,j}$ are functions of column pressure P, which then in turn determine the vapor composition $y_{i,j}$. This model analysis indicates that the pressure in the distillation column directly affects the mass balance and indirectly affects the energy balance (Eq.(7a)-Eq.(8b)) for all types of equilibrium model. The non-linear terms in the balance equations are related to the constitutive variables, which are associated to the process variables T, P, $x_{i,j}$ and $y_{i,j}$.

From a design point of view, the three design variables are selected first. The product rate D (or B) needs to be specified to meet the external mass balance and the market needs. Vapour is needed in the column to fulfil the separation process, while the heat needed to produce the vapour comes from specifying the heat input to the reboiler. A third process design variable that needs to be specified is the column pressure, because it affects the equilibrium of the system, and in turn determines the temperature level of the distillation column.

From a control point of view, for dual composition control of the distillation column, both the top product purity and the bottom product purity need to be controlled. For example the optimisation variable or design variable D (distillate flow rate) is used to control top composition x_D and vapour flow rate at the bottom V_B is used to control bottom purity x_B, where vapour flow rate is manipulated by heat duty of the reboiler Q_B. The column pressure is controlled by heat removed from the condenser Q_C. Hence for control, the design optimisation variables, Q_B, Q_C, D, may be chosen as the manipulated variables. This illustrates the relation between design and control issue for a conventional distillation column.

During the design of the distillation column utilities should be designed such that the column should be able to operate within some reasonable margins available for disturbance rejection. Limits on operation parameters form the borders of an operating window for the column. For the conventional distillation column an upper limit of the vapour flow rate is defined by the flooding limit and a lower limit by the weeping limit, these may be illustrated in an operation window plot displaying V as a function of P, for a specific D.

Model Analysis for the Energy Integrated Distillation Column

The process model of the integrated column can be represented by Eq(1)-Eq(15). The specified variables are still three. The difference is that two of the design variables or optimisation variables (the heat addition Qc and removal term Q_B) of the integrated distillation column are now a function of the process variables in the heat pump (see Eq.(12b)-Eq(15)). From a design point of view, a more energy efficient process is obtained, but from a control point of view, the optimisation variables Q_C and Q_B can't be manipulated directly, since they are functions of the properties of the heat pump fluid. So, before investigating the design and control structure of the energy distillation column, the properties of the heat pump fluid relative to those of the distillation components are examined.

PROPERTIES OF THE PURE COMPONENTS

Figure 3 shows plots of pure components heats of vaporisation and vapour pressure for heat pump fluid (freon-114) and the distillation components methanol, iso-propanol and water. From Figure 3 we can see that on the heat pump side, disturbances in the heat exchangers affect the *PVT* (pressure-volume-temperature) behaviours of the pure component. Therefore, we are at this stage only concerned with the heat of vaporization and vapour pressure for the pure components as a function of temperature for the heat pump fluid, methanol, isopropanol and water.

Qualitative Analysis from the Diagram of Pure Components

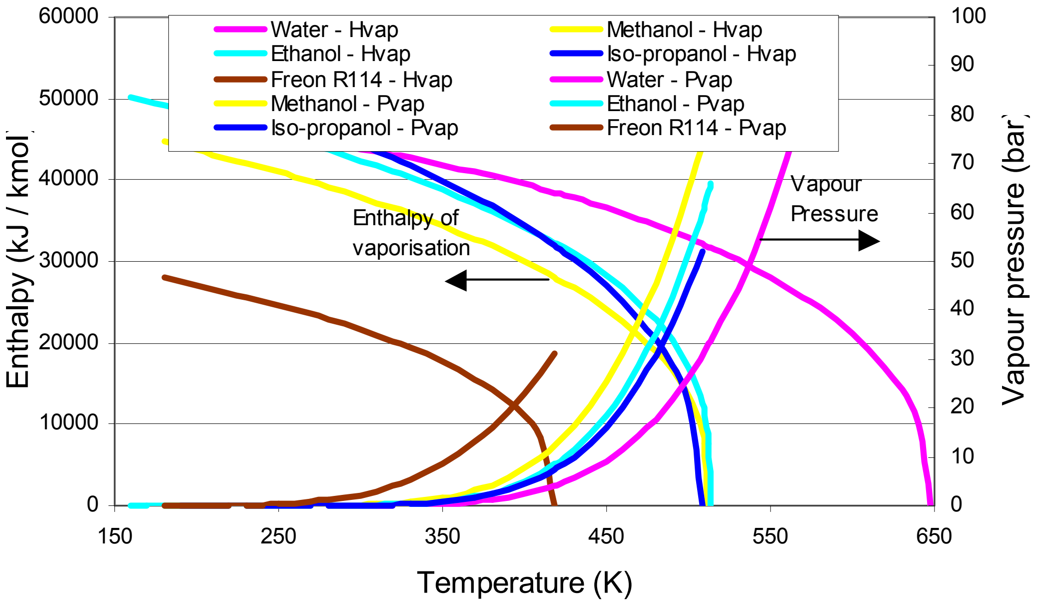

Figure 3: Heat of vaporization and vapour pressure as functions of temperature for refrigerant and column components.

We first focus on the enthalpy and vapour pressure curves of the freon-114. In Figure 3 both the pressure and the enthalpy curves are non-linear and calculated through the property prediction package in ICAS (Gani et al.(1997)[8]. At high temperature or high-pressure area the slope of these two curves are steeper. Thus one can imagine that a small change on the heat pump side in pressure may cause large changes in temperature on this side. Consequently, a large change in the vapour boil-up on the column side is produced since the change in temperature causes a change in vaporisation. A similar analysis for the condenser on the distillation column side and heat pump side shows that temperature on the heat pump side is sensitive to small changes in pressure, but the reflux flow rate is less sensitive to a change in temperature, i.e., a change in temperature may cause a change in condensation, but since the residence time for a liquid is generally larger than that of vapour, the liquid (reflux) flow rate is less sensitive than the vapour (boil-up) flow rate. Thus, the plots of the constitutive variables (such as heats of vaporisation) as functions of the process variables (such as temperature) provide an indication of the feasible range of operation for design and control. Based on this analysis, the pressure on the heat pump side needs to be controlled at both ends in order to ensure reachability of a specific static operating point for the total process. Pilot plant operation and controller design for our integrated distillation column (see Figure 1) confirm this analysis where the high pressure P_H is controlled by valve CV8 and the lower pressure P_L is controlled by valve CV9 on the heat pump side. The operation window will be selected at the lower pressure on the condition that it can satisfy the operation requirement. The quantitative relationship between both sides is useful in providing the sensitivities of the integrated distillation column and they will be discussed below.

Quantitative Analysis

The quantitative relationship is analysed with a simulation program from Koggersbøl[9]. The basic control configuration for the plant for these simulations is as follows: The high pressure P_H is controlled by CV8, the low pressure P_L is controlled by CV9, accumulator level is controlled by L_o, and reboiler level is controlled by B. Q_B and Q_C are then manipulated through the setpoints to P_H and P_L since these indirectly affect the temperature differences in the reboiler and the condenser. First, the operation condition is changed, i.e. P_H is increased from 1100 to 1600kPa with in steps of 2kPa to observe the influence of this disturbance on pressure and composition of the column. Some results are plotted on Figure 4 and Figure 5, where Figure 4 shows the changes of the column pressure including the bottom pressure and top pressure to this disturbance and Figure 5 shows the composition changes to this disturbance ($x_{1,D}$ is the methanol composition on the top, $x_{2,B}$ is the isopropanol composition at the bottom) and operation points move from point 1 to 8. Figure 4 and Figure 5 show that for the same disturbance of P_H on the heat pump side, the effect on the column pressure and composition to this disturbance is different and the changes increase while the operation condition P_H increase. For example, if P_H is set at 1100kPa, then for a 2kPa disturbance introduced, the bottom pressure difference increases to 269.9 Pa and if P_H is controlled at 1600kPa, then for a 2kPa disturbance introduced, the bottom pressure difference increases to 516.3Pa. When the operation points are moved from point 1 to 8, the top methanol composition changes from 0.97549 to 0.9671 and the bottom isopropanol composition change from 0.9637 to 0.9567(molar fraction). The pressure and composition difference in sensitivity confirm the properties of the pure components, because at high pressure the slope of the pressure and enthalpy with temperature is steeper. The column system operation is more sensitive at high-pressure.

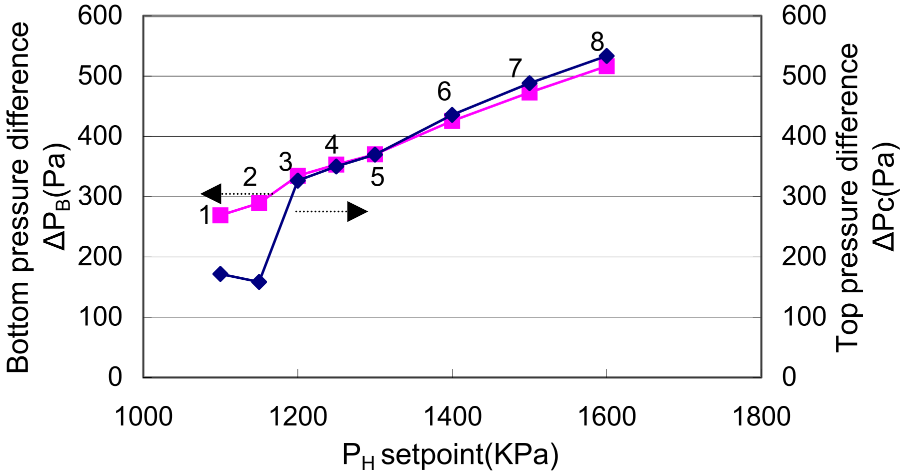

Figure 4: The column pressure's sensitivity to ΔP_H=2kPpa at different high pressure set-points: P_H=1100, 1150, 1200,1250, 1300, 1400,1500,1600(kPa) D=1545.89mol/h

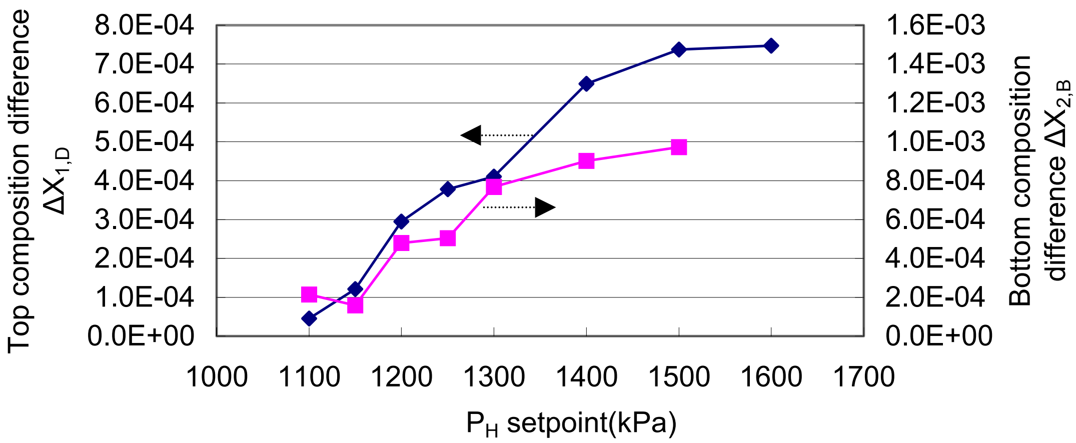

Figure 5: The column composition's sensitivity when ΔP_H=2kPa as a step at different high pressure set-point P_H=1100, 1150, 1200, 1250, 1300, 1400,1500,1600(kPa), D=1545.89mol/h
$X_{1,D}$ is the methanol composition on the top
$X_{2,B}$ is the isopropanol composition at the bottom

OPERATION WINDOW OF THE ENERGY INTEGRATED DISTILLATION

In order to investigate the design and control issue of the integrated distillation column, the operation window needs to be identified. For a given sieve-tray design and column pressure, there exists a minimum vapour flow rate below which the vapour cannot carry the liquid on the tray and liquid begins to weep through the holes. Correspondingly there exists an upper limit for the vapour flow rate above which the vapour stream carries some of the liquid to the tray above. This phenomenon is called liquid entrainment and the limit is usually specified by the fraction of the liquid flow rate that is entrained. At an even higher vapour flow rate the flow pattern on the tray may break down because the down flowing liquid cannot overcome the pressure drop through the tray. The tray then starts to flood and no separation takes place anymore. At very high liquid loads some vapour may be captured as bubbles in the down flowing liquid and this may be returned to the tray below. This is called vapour entrainment.

In the case of our heat integrated column with sieve trays operated primarily in the spraying regime, the limits forming the operating region are the flooding limit, weeping limits, maximum column pressure, maximum heat pump high pressure, maximum heat pump low pressure, maximum cooling power of the heat pump system, maximum pumping capacity. Inside this region is the operation window, within which the operation point must be located to ensure the separation process. Figure 6[5] is one kind or operation window, which contains weeping limit, flooding limit, maximum low pressure, and minimum low pressure. These low and high pressure limits are determined by the selected refrigerant combined with their equipment design. However these limits are not the key topics of this paper in stead Koggersbøl (1995)[9] may be committed. In this paper the operation within the boundary defined by these limits is analysed.

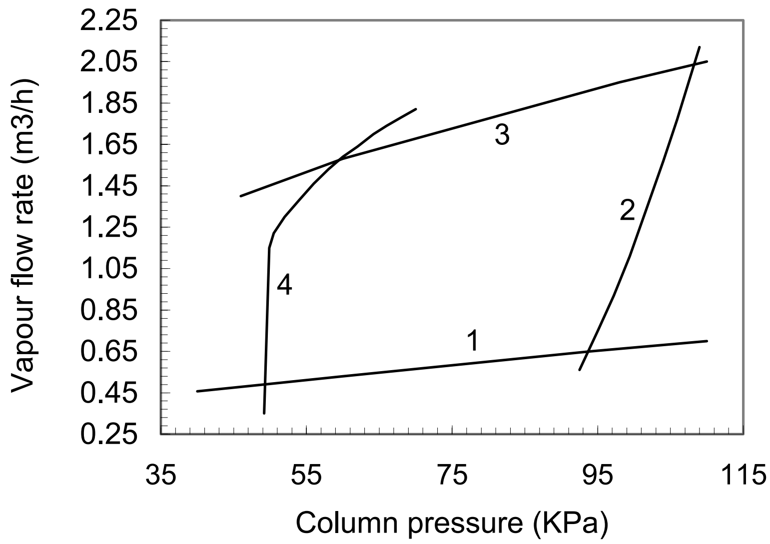

Figure 6: Operation window

1.Weeping limit;

2. Maximum P_L

3. Flooding limit;

4. Minimum P_L

THE DESIGN AND CONTROL OF THE INTEGRATED DISTILLATION COLUMN

Analysis of the Movement of the Operation Point

For the energy integrated distillation column, the production rate D (or B) will still be designed to meet the desired amount. The vapour flow rate and column pressure in the distillation column also need to be designed for the separation process. But the heat pump is not the actuator of Q_B, Q_C. Hence the first question is which heat pump design variables to select and second how to manipulate vapour flow rate in the column.

First, simulations are done as listed in Table 1, where the heat pump low pressure P_L is kept constant at 500kPa, P_H is decreased in steps of 25kPa. The column pressure (top tray) and the boil-up flow rate are given in Table 1. For these simulations the feed composition of methanol X_F is 0.4950(mole fraction); X_D (top composition of methanol) and X_B(bottom composition of isopropanol) are 0.9730 and 0.9609 respectively.

Then, the heat pump high pressure P_H is kept constant at 1075kPa; P_L is decreased in steps of 25kPa. The simulation results list in Table 2.

These results are plotted in Figure 7, where branch A is the result in Table 1 and branch B is the result in Table 2.

Table 1 (Branch A)

Point	1	2	3	4	5
P_H(kPa)	1175	1150	1125	1100	1075
P_L(kPa)	500	500	500	500	500
P_C(kPa)	81.2196	78.9842	76.5508	73.8749	70.8727
V(m^3/h)	1.423	1.184	0.9497	0.720	0.504

Table 2 (Branch B)

Point	5	6	7	8	9
P_H(kPa)	1075	1075	1075	1075	1075
P_L(kPa)	500	475	450	425	400
P_C(kPa)	70.8727	69.9443	68.3358	66.3793	64.1274
V(m^3/h)	0.504	0.782	1.040	1.286	1.516

From Figure 7 one can see that either manipulating constant heat pump high pressure P_H or manipulating constant heat pump low pressure P_L cannot control the column pressure and boil-up flow rate. As decreasing P_H at constant P_L the operation points move along curve A from operation point 1 to 5. When decreasing P_L at constant P_H, the operation points move along curve B from operation point 5 to 9. Both curve A and curve B are nonlinear and sensitive to the pressure change of the heat pump side. So this control structure will not be able to satisfy the desired operation requirement, which is to maintain constant column pressure and vapor flow rate on the column.

Analysis the Design and Control Structure
As discussed above, the two optimisation variables (Q_B, Q_C) are shifted to the heat pump side in the integrated distillation column. Hence the column pressure and vapour flow rate must be controlled by manipulating P_H and P_L. However direct manipulation renders it difficult, due to large changes in sensitivity within the operating window as shown in Figure 7. To analyse this behaviour the process model is useful. If for instance, P_H is increased at constant P_L this will affect first the temperature difference between both sides in the reboiler ($\Delta T_{Reboiler}$). This causes the vapour flow rate to increase and so does the pressure drop ΔP. To accommodate the higher vapour flow rate the pressure in the column of top increases until the temperature difference in the condenser satisfies the heat balance. The high pressure actuator, thus, has positive gain to both boiler-up rate and column pressure. If instead P_L is increased at constant P_H the temperature difference between both

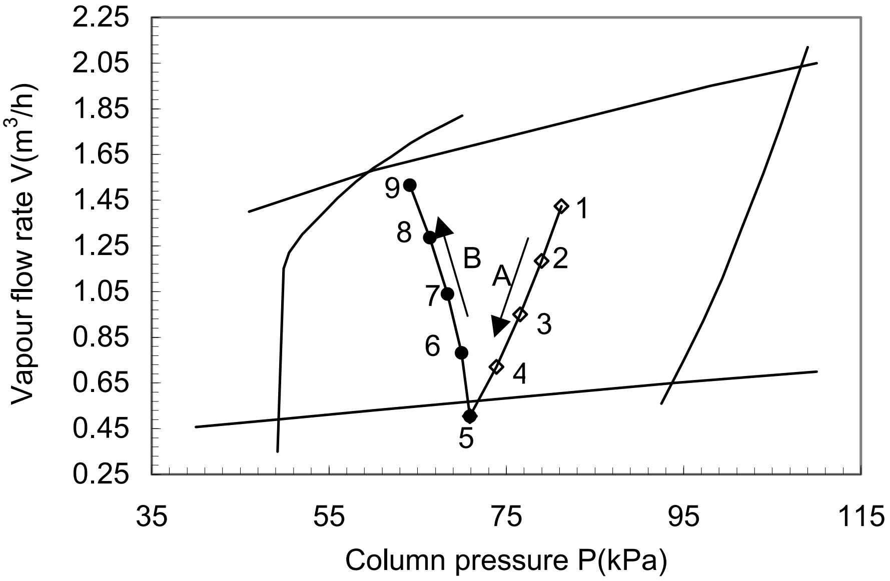

Figure 7: Operation points movement (measured on the top tray) and boil-up flow rate.
Branch A: Constant P_L=500kPa, 25Kpa steps decrease in P_H
Branch B: Constant P_H=1075kPa, 25 KPa steps decrease in P_L.

sides in the condenser ($\Delta T_{Condenser}$) is reduced. Thereby the column is elevated to a high pressure but at the same time the smaller ΔT results in reduced boil-up rate. The low pressure actuator, thus, has positive gain to the column pressure but negative gain to the boil-up rate. The gains for positive and negative changes in the heat pump pressures are illustrated on Figure 8.

From this understanding it is seen that in order to increase the column pressure at constant boil-up rate one must increase both actuators, while if the boil-up rate is to be increased at constant column pressure (either P_B or P_C) one must increase P_H and reduce P_L. So it is clear that specifying the two heat pump pressures P_H and P_L is equivalent to specifying boil-up flow rate and column pressure and hence it should be possible to configure a control system manipulating the setpoints to the high and low pressure (or perhaps the control valves CV8 and CV9, see Figure 1) directly such that the operator can manoeuvre the process through the operating region using the actuators P and V of an conventional distillation column. The selection of the above actuators (P_H+P_L) for column pressure and (P_H-P_L) for vapour flow rate ensures a simple mapping to the conventional actuators P, V. In other words, with this basic control configuration the column can be viewed as a "standard" distillation column with the following actuators: L_0, V, D, B, and column pressure. So the static control and design structure of the integrated distillation column are clear, even when controlled by the high and low pressure on the heat pump side.

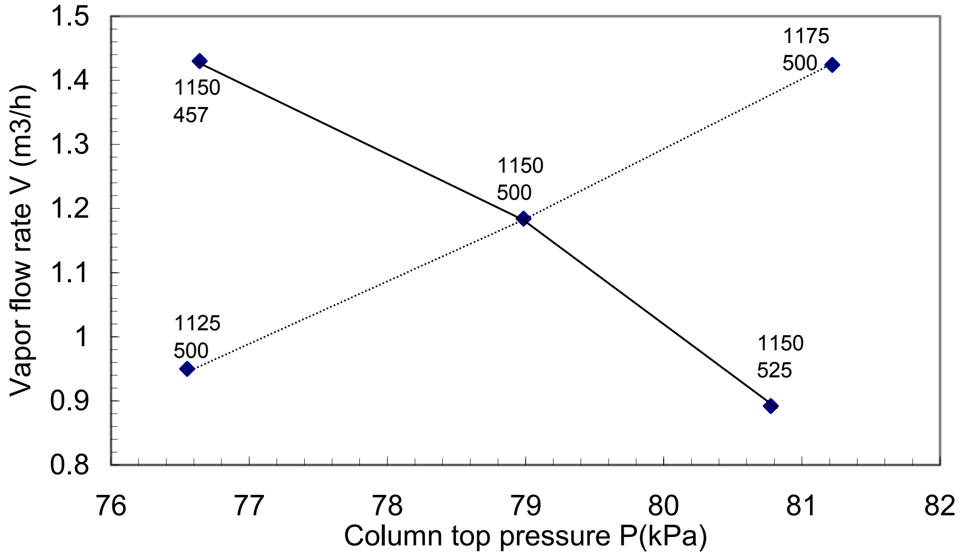

Figure 8: The gain for positive and negative changes in the heat pump pressures
Steps $\pm 25kPa$ in high and low heat pump pressures
(on simulation plot: P_H over P_L in kPa)

Actuator Configuration

To confirm the design and control issues discussed above, dynamic simulations have been performed with the dynamic simulation program of Koggerbøl [9]. First, P_H+P_L is set as constant, while P_H-P_L is allowed to change and then P_H-P_L is set as constant, while P_H+P_L is allowed to change. The simulation results are plotted in Figure 9, where the curve A in Figure 9 is at constant P_H+P_L, while P_H-P_L change. From curve A one can see that column pressure is nearly constant for constant P_H+P_L for many different P_H-P_L. From curve B one can see that the vapour flow rate is nearly the same at constant P_H-P_L in spite of different P_H+P_L. This confirms the design and control issues stated above that one could control the vapour flow rate in the integrated distillation column by manipulating the pressure difference between the high pressure and the low pressure of the heat pump side and the column pressure by manipulating the sum of the two heat pump pressures.

The above analysis shows that design and control aspects of the integrated distillation column are closely related. These relationships have been illustrated above; several additional relations can be uncovered from analysis of the pure component properties.

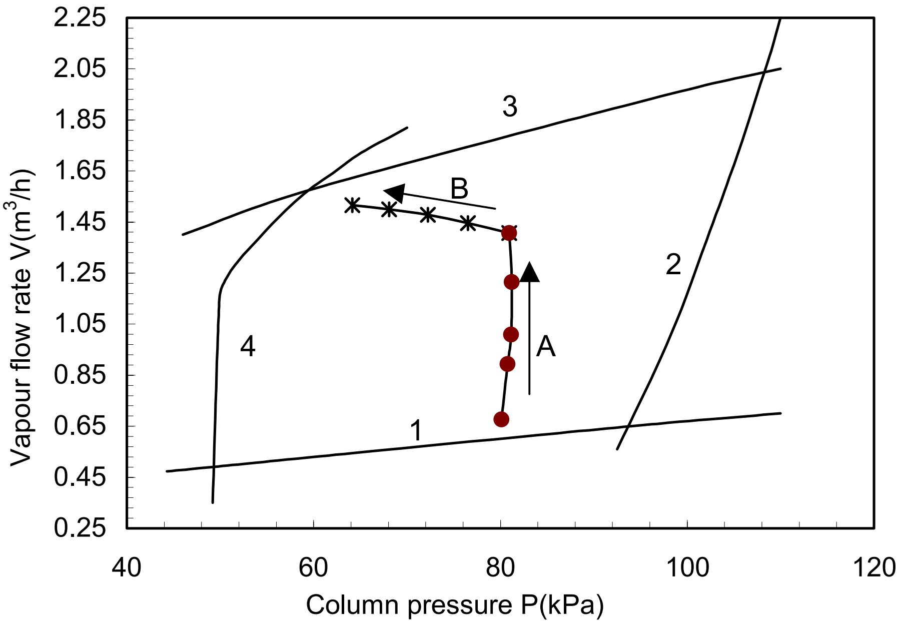

Figure 9: A curve: P_H+P_L at constant, P_H-P_L change
B curve: P_H-P_L at constant, P_H+P_L change

CONCLUSION

Computer aided model analysis provides useful information related to the formulation and solution of integrated design and control problems. This information helps to identify the important process and design variables. The design and control actuator selection for an integrated distillation column is carried out through a static model analysis. Rigorous simulation results verify this analysis. This type of analysis holds the promise that an integrated design and control problem can be defined correctly and consistently with respect to the process behaviours and the selected process models. The key contribution of this work is the qualitative analysis and insights gained from the model analysis. It is demonstrated that determination of a suitable control structure can be obtained for a multivariable problem. Thereby this paper demonstrated an analytical method for obtaining insight into gain directions for multivariable control problems.

NOMENCLATURE

C_p	Heat capacity of refrigerant
B	Bottom product flow rate
D	Top product flow rate
F_j	Feed flow rate
H_D^L	Enthalpy at the condenser on the column side
H_B^V	Enthalpy at the reboiler on the column side
H_j^F	Enthalpy of liquid on feed tray
H_j^v	Enthalpy of vapour on tray j
H_j^L	Enthalpy of liquid on tray j
$H^H{}_{vap}$	Heat of vaporization of freon
$K_{i,j}$	Equilibrium constant of component i on tray j
L_0	Reflux flow rate
L_j	Liquid flow rate from tray j
$L^H{}_{EXP}$	The flow rate of refrigerant on heat pump side
NC	The total number of components
NP	The total number of trays
P_c	Pressure at the top
P_B	Pressure on the bottom
P_T	Pressure on the top
P_H	High pressure on the heat pump side
P_L	Low pressure on the heat pump side
ΔP	Pressure drop on each tray
ΔP_C	Pressure difference at the top
Q	Heat exchange before compressor
Q_j	Heat added or removed from tray j
Q_c	Heat removed from the condenser
Q_B	Heat add from the reboiler
t	Time
T	Temperature
$\Delta T_{Reboiler}$	Temperature difference between both sides in the reboiler
$\Delta T_{Condenser}$	Temperature difference between both sides in the condenser
T_j	Temperature on the tray j
T_3	Inlet temperature of expansion valve
T_6	Inlet temperature of the compressor
V_B	Vapour flow rate at the bottom of the column
V_1	Vapour flow rate on top tray
V_j	Vapour flow rate on tray j

$X_{i,j}$	Mole fraction (liquid) of component i on tray j
$X_{1,D}$	Methanol composition on the top of the column
$X_{2,B}$	Isopropanol composition at the bottom of the column
$Y_{i,j}$	Mole fraction (vapour) of component i on tray j
$Z_{i,j}$	Feed composition

SUBSCRIPTS

c	Top
i	Component i
j	Tray j
k	Component k

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