[Proceeding] On the optimisation and control of Pressure Swing Distillation unit

On the optimisation and control of Pressure Swing Distillation Unit

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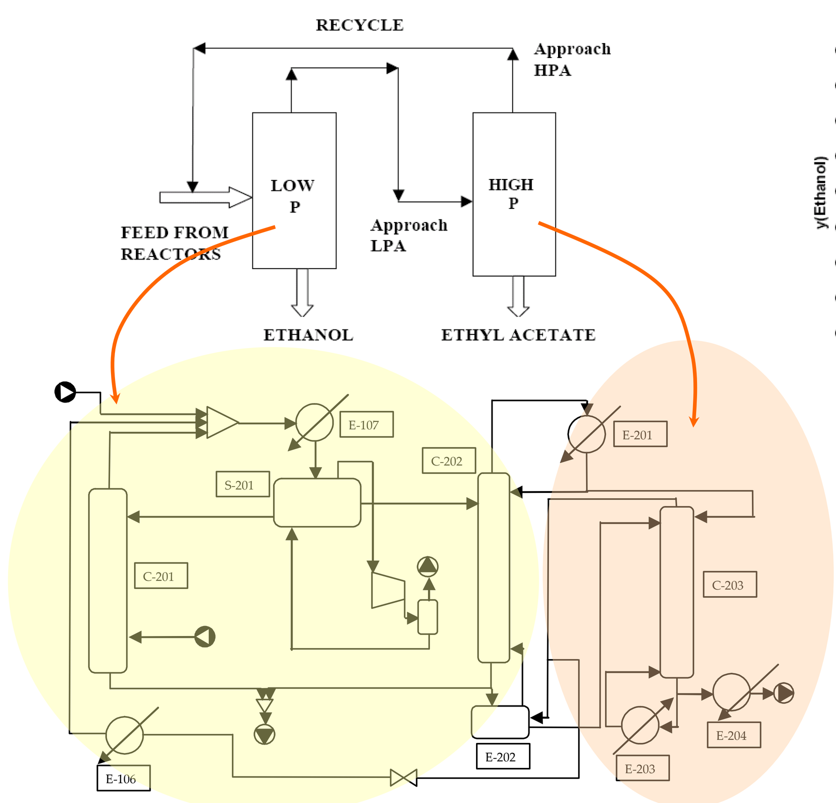
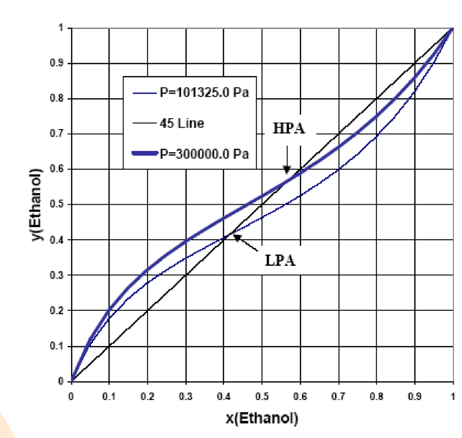
Introduction

Recycling of process solvents is receiving a good deal of attention lately as engineers and chemists strive to reduce long-term costs and minimize wastes. Distillation can be used for handling these recycle streams because it is a familiar and robust unit operation and because the required unit equipment may already exist on-site.

The presence of azeotropes, however, may severely limit the use of standard distillation. This work deals with the modelling of a **Pressure Swing Distillation (PSD)** Unit for the recovery of ethyl-acetate from a mixture containing also ethanol, water and other organic compounds in small amounts, being these components responsible for the formation of pressure-sensitive azeotropes and of liquid-liquid splits; moreover in a PSD unit the sequence of the columns can be readily thermally integrated, thus allowing further savings.

Pressure Swing Distillation concept





The control problem

Constraints: Ethyl-acetate recovery > 90 %; Purity of the ethyl-acetate recovered > 99.5 %; Water concentration in the product < 0.1 %; Energy integration between the low pressure and the high pressure column.

The Disturbances: Feed flow rate and composition are variable with time; Pressure in the decanter and in the low pressure column may be variable.

The objective: Minimise the cost of the operation, i.e. the duty in the high pressure column

Model Predictive Control is the usual framework for optimising plant performances, but the number of the optimised variables may render the on-line optimisation particularly difficult, thus requiring the use of simplified model (even black-box model).

Because of the complexity of the system, a **detailed model should be used**; this choice allows also to use the MPC algorithm on plants of different size or different operating conditions, without the need of re-evaluating the parameters of the (simplified or black-box) model.

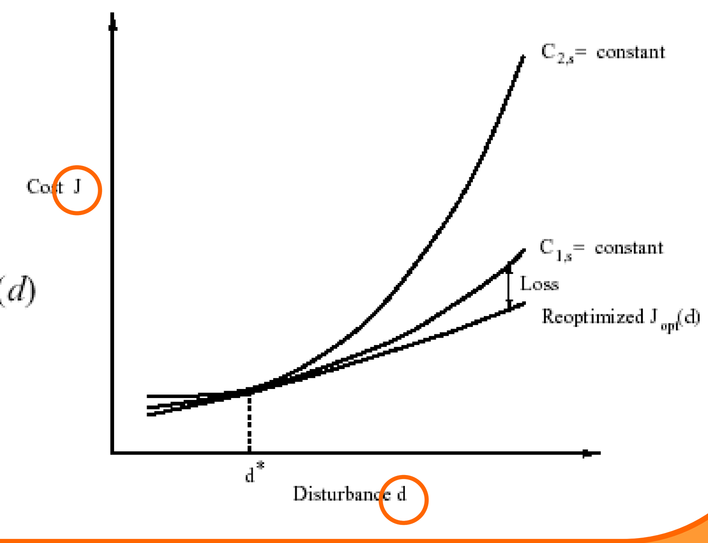
Process optimisation

The main drawback of working with a detailed model in a MPC algorithm is the **complexity of the on-line optimisation**; as a consequence steady-state optimisation was used to point out the influence on the objective function of the variables that can be manipulated, showing that the optimisation of one variable properly chosen may be enough, being poor the effect of the others on the plant optimisation.

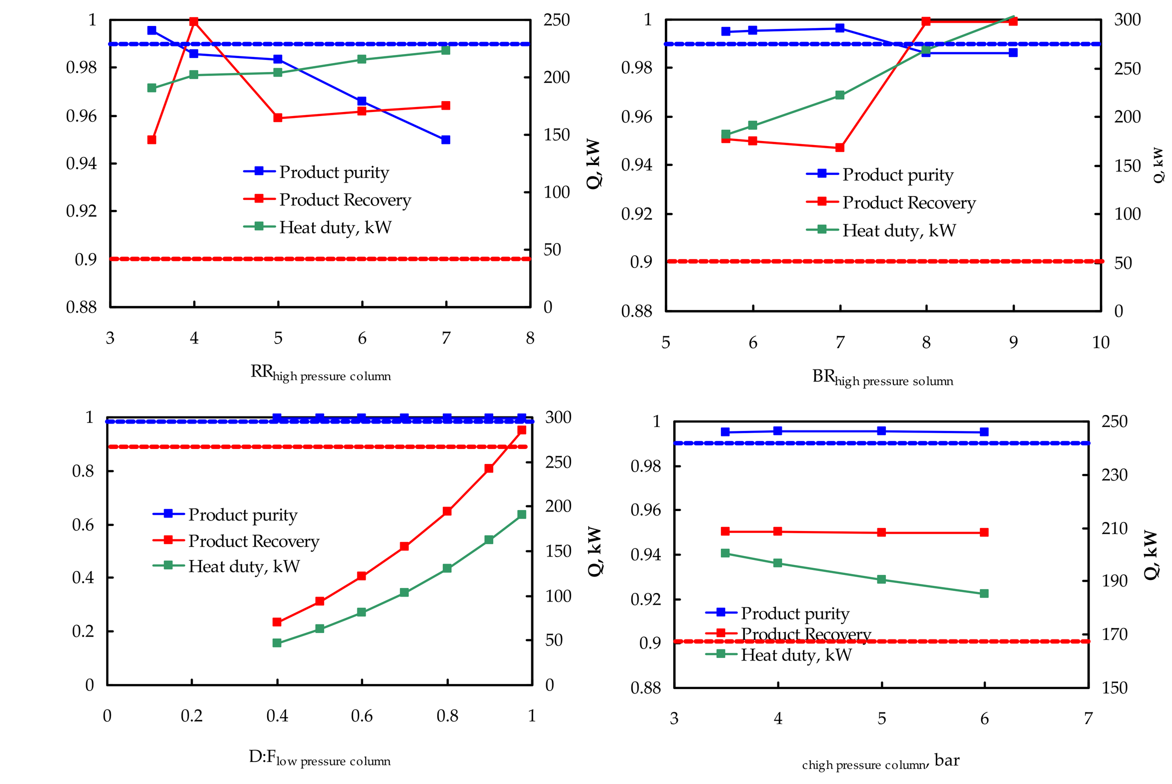
$$\begin{cases} \min\limits_{u} J(u,d) \\ g(u,d) \le 0 \end{cases}$$

$$\min_{\substack{u \\ g(u,d)\le 0}} J_u(u,d) = J_u\left(u_{opt}(d),d\right) = J_{opt}(d)$$

$$L = L_u(u,d) = J_u(u,d) - J_{opt}(d)$$

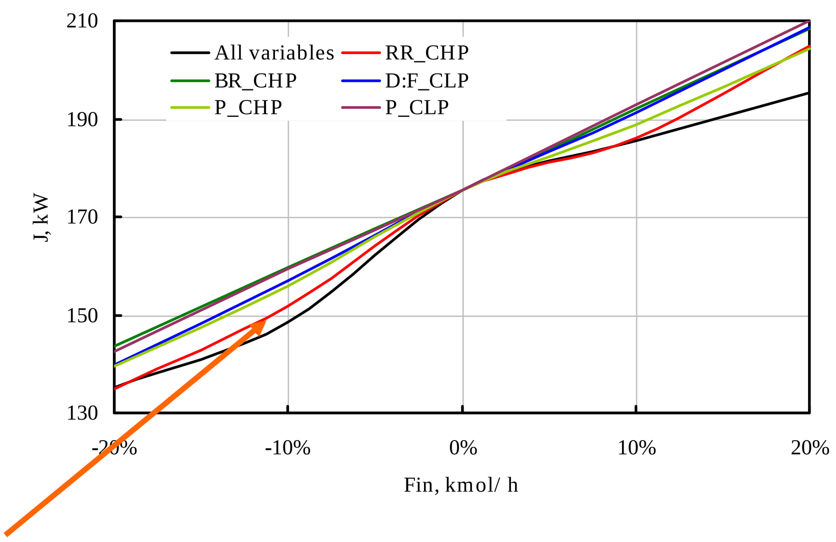


Sensitivity analysis



Process optimisation

In order to point out the role of the various manipulated variables on the cost function we compare the results obtained when all the variables are optimised and when just one is optimisated in presence of various disturbances.



The optimisation of the reflux ratio of the high pressure column is enough to fulfill the constraints and to minimise the cost function

Model Predictive Control Algorithm

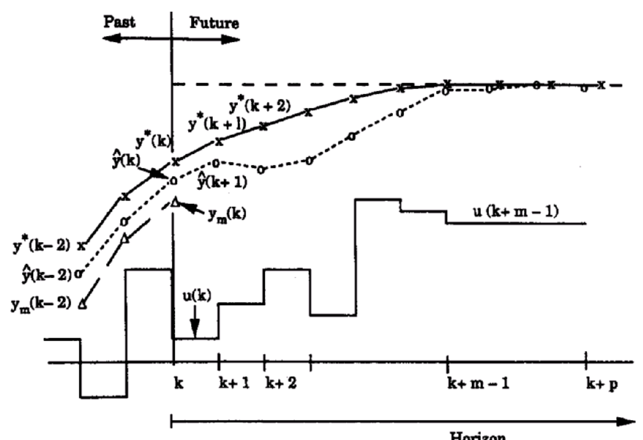
$$\min_{\hat{u}}\left(\sum_{i=1}^{h_p}\left(y(i)-y_{set}(i)\right)^2\right)$$

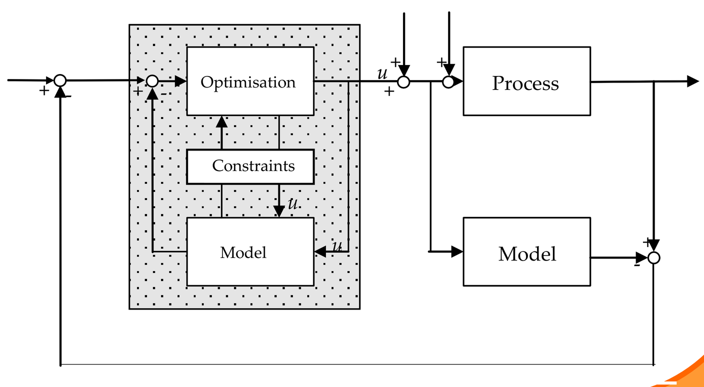
$$\min_{\hat{u}(k)\ldots\hat{u}(k+h_c-1)}\left\{\sum_{j=k+1}^{k+h_p}\omega_y(\hat{e}_y(j))^2+\sum_{i=k}^{k+h_p-1}\omega_u(\Delta\hat{u}(i))^2\right\}$$

Constraints handling: *penalty functions*

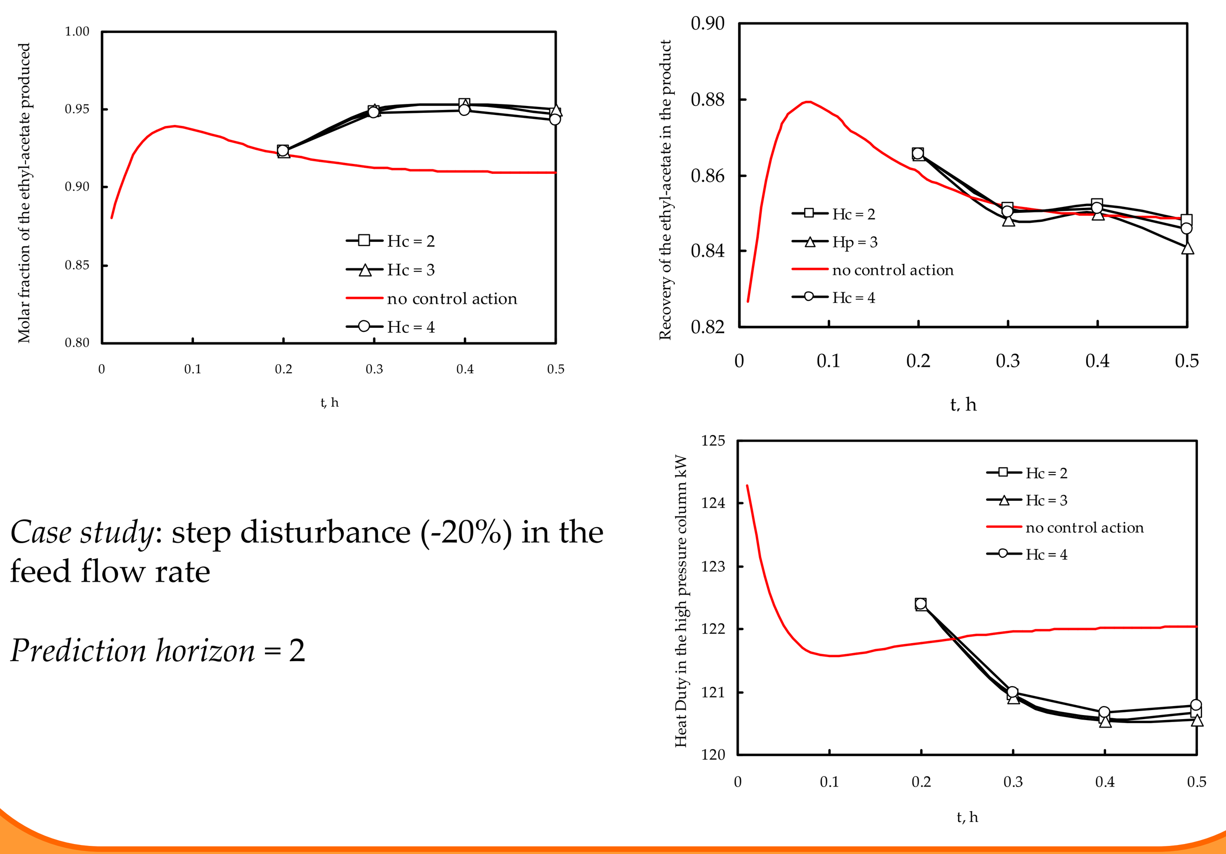
$$F = F + \sum_{t=k+1}^{k+h_p}\left\{\sum_{i=1}^{nc}\omega_{i,SOFT}\left[\frac{y_i(t)-y_{i,SOFT}}{y_{i,SOFT}}\right]^2\right\}$$

Model error handling:





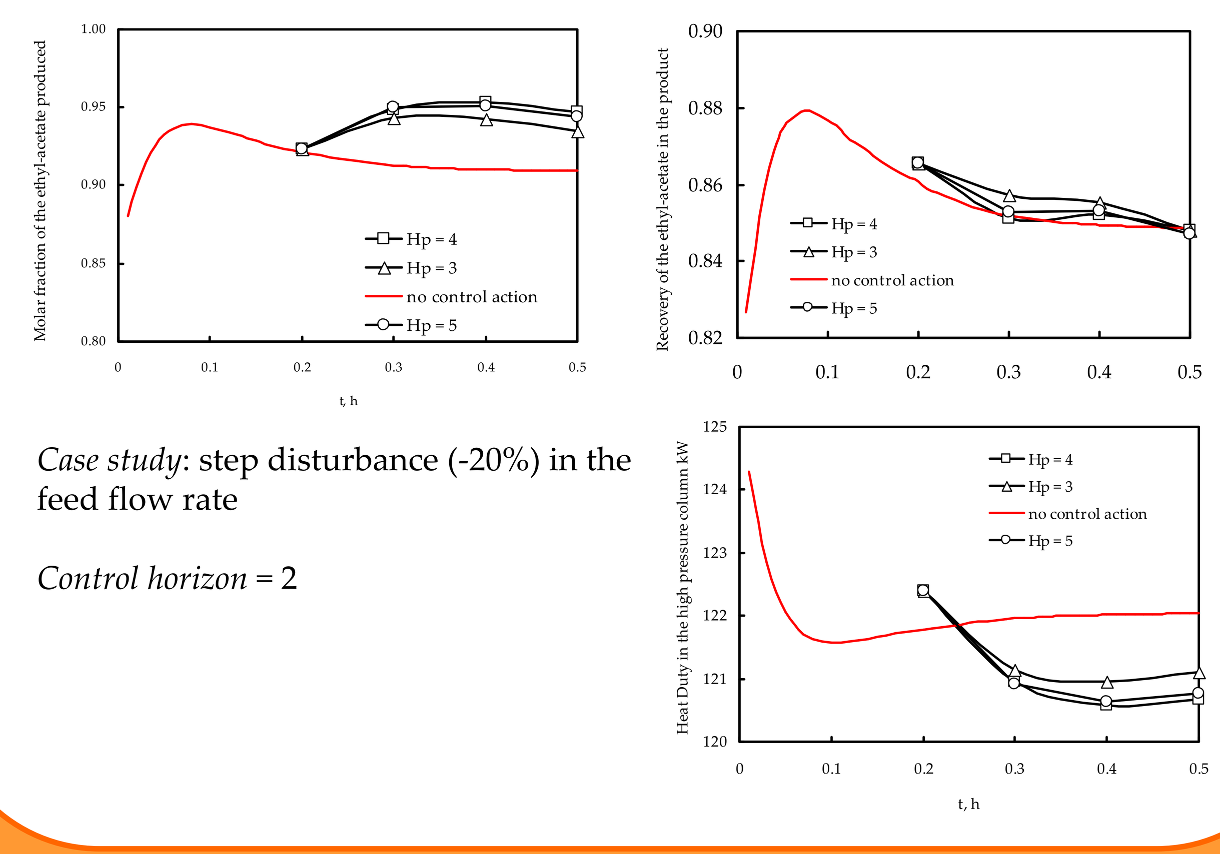
Influence of the parameters of the MPC algorithm



Case study: step disturbance (-20%) in the feed flow rate

Prediction horizon = 2

Influence of the parameters of the MPC algorithm



Case study: step disturbance (-20%) in the feed flow rate

Control horizon = 2