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MSF design taken into account availability

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Abstract

In this paper a methodology for MSF process optimization taken into account process availability will be presented. The process optimal operative conditions, and the optimal equipment allocation in stand-by units when necessary will be performed, given a process model and cost data.

Keywords: MSF desalination processes; Availability; Maintenance; MINLP model; Optimal synthesis and design

1. Introduction

It is well known that maintenance and reliability play an important role in process operability and productivity. Availability depends strongly of the probability of equipment failures, which in turn depends on the engineering design and the operation environment. However, without maintenance intervention, equipment will ultimately fail, and result unavailable. So, the mean time an equipment works (is reliable) and the time employed to repair it are very important economic factors. In fact, the number of stand by units, the number of operation and maintenance personal needed, greatly influence in the operational and investment capital cost.

Even it is accepted that a focus on availability is critical to optimize plant operation and productivity, no sufficient effort was made to introduce this point of view in process optimization. In order to consider availability and maintainability data in process models, cost functions describing relationships among capital investment, maintainability and reliability unit costs, must be introduced. So, the problem is how to determine the optimal parameters, and at the same time, the optimal reliability and maintainability for each process unit that maximizes the annualized expected profit (taken into consideration revenues, equipments costs, operating and maintenance costs) or minimize the expected production costs, if this is the case [1–13].

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will be presented. The process optimal operative conditions, and the optimal equipment allocation in stand-by units when necessary will be performed, given a process model and cost data. It will be shown that the problem to be solved is a MINLP of the form:

$$\text{Min } \Phi(x, y, \text{As})$$

$$\text{st. } h(x, y) = 0$$

$$g(x, y) \leq 0$$

$$A_{\min} \leq \text{As} \leq 1$$

$$y \text{ € } Y \; [0, 1]$$

$$x \text{ € } X R^n$$

where $\Phi(x, y, \text{As})$ is a scalar economic objective function (total annual cost), x is a vector of continuous variables specified in the compact set X. Equality constraints correspond to the process model, covering mass and energy balances, and physical-chemical relationships, while $g(x, y) \leq 0$ is a vector of inequality constraints corresponding to process design and operative restrictions. Finally, As is a scalar continuous variable describing process availability, which in turn is a function of the reliability (R) and the maintainability (M) vectors. Each vector contemplates the corresponding variables for each process unit. Finally, $A_{\min}$ is a minimum value for the process availability imposed by the designer.

In general, the achievement of more process availability is often obtained by equipment redundancy using series or parallel configurations (or more complex arranges); but also each equipment availability can be improved by expending more investment capital in its reliability and or maintainability. So, not only the basic process variables must be considered, but also the system reliability configuration (redundancy structure for example). In fact, while the first is influenced by the physical arrange of equipments and their interconnections, the system availability As depends on the logic underlying the units allocation and its individual reliability and maintainability.

Obviously, if we want to permit economic trade-off among the extra capital charge and operative costs due to the objective to achieve more reliable and/or maintainable equipment, or to acquire better configurations to improve system availability; we must introduce these variables in the problem formulation [3,14]. To show how to solve the above problem, in section 2 we present the process model, in section 3 we will define our cost model to define our objective function, in section 4 we will analyze the reliability and maintainability models, and finally, in section 5 a study case will be presented.

2. Process model

In this case, a multi-stage flash mixer unit will be analyzed. Of course, this methodology is also valid for other systems. Fig. 1 briefly illustrates the process. The seawater feed (SW) previously treated with antiscalant chemicals after screening is mixed with the recycle stream (WR) and it is heated as it flows in series in the condenser tubes at each stage and the brine heater, being heated to a temperature greater than the saturation temperature at the final stage. It then flows into the first stage inlet box (flash chamber) and it is evenly distributed across the width of the evaporator. It enters each stage through an orifice and a weir system that control its flow rate and the flashing characteristics. Then, the flashed vapor flows first through a

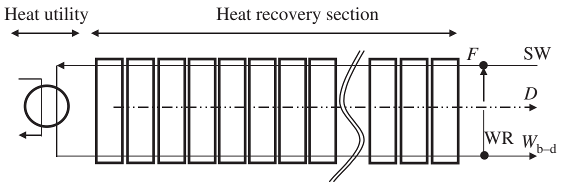

Fig. 1. Multi stage flash — mixer desalination system.

demister, which removes any entrained brine, and then over a condenser that condenses it. The vapor release velocity from the flashing brine surfaces shall be as low as possible to minimize brine carry-over. The maximum vapor release velocity shall not exceed 8–9 m/s and is ensured by providing an adequate cross sectional area (length and width) for the flashing chamber.

The condensate, referred to as distillate (D), is collected in a distillate collector. The seawater, now referred to as brine since its salinity has increased, then flows through the evaporator stages in turn, releasing flashed vapor at each stage in the same way. The brine ($W_{\text{b–d}}$) is rejected from the last stage evaporator by a brine blow down pump to the sea. Part of this stream is recycled (WR) and mixed with the feed (SW) to enter the pre-heater tubes as previously described (F).

The distillate accumulated in each stage passes through the transfer system into the next lower temperature stage, where a proportion is similarly flashed to the brine. This vapor flows over the stage condenser together with the vapor flashed from the brine, and it is condensed and transferred to the next stage. The distillate accumulates as it flows through stages and is discharged from the last stage to a product water tank by the distillate pump. According to this description, the following equations can represent the process behavior adequately [15].

$$Q^{\text{Des}} = F\,C_{\text{p}}\,\Delta t \tag{1}$$

$$\Delta t = \Delta t_{\text{f}} + \Delta t_{\text{e}} + \text{BPE} \tag{2}$$

where Q^{Des} is the heat consumption, and Δt_{f}, Δt_{e} and BPE are temperature drop for the flashing operation, effective driving force for the heat transfer operation and boiling point elevation, respectively. The total heat transfer area A_{t} can be calculated by:

$$A_{\text{t}} = (\text{F}C_{\text{p}}/U)\ NS\ \ln\left((\Delta t - EPE)/\Delta t_{\text{e}}\right)^{(T_{\max}-\Delta t-T_0)/\Delta t_f} \tag{3}$$

NS (the number of stages), $T_{\max}$ and T_0 (the top brine and the sea-water temperatures, respectively) are related by the following equation:

$$\text{NS}\,\Delta t_{\text{f}} = T_{\max} - \Delta t - T_0 \tag{4}$$

The total production of distillate to be produced in the operative available time (Prod), is:

$$\text{Prod} = F\left[1 - \left(1 - \frac{C_{\text{p,m}}\Delta t_{\text{f}}}{\lambda}\right)^{\text{NS}}\right] \tag{5}$$

The relation among the heat transfer area (A_{t}) with the number of tubes (N_{t}) and the chamber width (B) and height (HS) are:

$$A_{\text{t}} = \pi \text{TD} B N_{\text{t}}\ \text{NS} \tag{6}$$

$$\text{HS} \cong 2\text{LB} + D_{\text{S}} \tag{7}$$

The following are relationships between the number of rows of tubes in the vertical direction (N) and the number of tubes N_{t}, among the shell diameter (D_{S}), number of rows of tubes and Pitch P_{t}, and the length of the desaltor [16]:

$$N = 0.481\sqrt{N_{\text{t}}} \tag{8}$$

$$D_{\text{S}} = N\ P_{\text{t}}\ \sqrt{2} \tag{9}$$

$$\text{LD} = \frac{\text{Prod}}{B\,V_{\text{vap}}\,\rho_{\text{vap}}} \tag{10}$$

where V_{vap}, ρ_{vap} are the velocity and density of the vapor respectively. The total stage surface area is:

$$A_{\text{S}} = 2\text{LB} + 2\text{HL} + H_{\text{S}}\,B \tag{11}$$

Finally, the following constraint must be satisfied:

$$\text{LD} = \text{D}_{\text{S}}\,\text{NS} \tag{12}$$

$$F = \mathrm{SW} + W_{\mathrm{R}} \tag{13}$$

$$F = \mathrm{Prod} + W_{\mathrm{R}} + W_{\mathrm{b-d}} \tag{14}$$

$$W_{\mathrm{R}} = R\, W_{\mathrm{b-d}} \tag{15}$$

$$F\, C_{\mathrm{P}}\, T_{\mathrm{F}} = W_{\mathrm{R}}\, C_{\mathrm{p}}\, T_{\mathrm{b-d}} + SW\, C_{\mathrm{p}} T_0 \tag{16}$$

where W_{R}, $W_{\mathrm{b-d}}$, and SW are the flow-rate values corresponding to recirculated, blown-down brine and sea-water streams. R is the recirculation ratio.

3. Objective function. Economic model

As was defined our problem in section one, here we minimize the total annualized cost for a given annual production (Prod), so:

$$\Phi(x, y, \mathrm{As}) = (\mathrm{TAC}) = M\, C_{\mathrm{op}} + C_{\mathrm{inv}} + C_{\mathrm{prem}} + C_{\mathrm{cmc}} \tag{17}$$

Where C_{inv} is the annualized investment cost, C_{cmc} is the corrective maintenance cost and C_{op} is the operating cost. C_{op} comprises the vapor cost in the heater, feed pretreatment costs and pumping costs. C_{inv} includes area investment costs (heat exchange area and total chamber area) and pump costs. For the stand by stages in the reliability block, the investment cost must consider the stand by units.

$$C_{\mathrm{inv}} = C_{\mathrm{A}}\, A_{\mathrm{to}} + \Sigma m_{\mathrm{s}}\, C_{\mathrm{ps}}\, Q_{\mathrm{s}}$$

where C_{A} is the unit area cost, A_{to} the total area, C_{ps} the cost per unit of flow-rate pumped, for each pump stage (s), and Q_{s} the flowrate in each stage s. As, according to our design restrictions in each pump stage we can allocate one, two or three pumps (two in stand-by), the investment cost in each stage is a function of an integer variable, handling the number of pumps in each stage, m_{s}. As explained, A_{to} is calculated summing the heat exchange area and the chambers area, affected by a factor Fact, which is known and represent the ratio of costs between these areas $A_{\mathrm{to}} = A_{\mathrm{t}} + \mathrm{Fact}\, A_{\mathrm{s}}$.

It is clear that investment costs are a function of the plant nominal capacity, but considering the process availability, are also function of the real capacity RProd, see Eq. (18). On the other hand, C_{op} is given by: $C_{\mathrm{op}} = C_{\mathrm{heat}} + C_{\mathrm{pump}} + C_{\mathrm{feed}}$, with C_{heat} the heat cost, C_{pump} the pumping cost and C_{feed} the pretreatment cost. Each term is calculated as: $C_{\mathrm{heat}} = C_{\mathrm{Q}}\, Q^{\mathrm{Des}}$ (C_{Q} is the unitary vapor cost and Q^{Des} is the heat consumption), and $C_{\mathrm{feed}} = C_{\mathrm{p}}\, F$, where C_{p} is the unitary pretreatment cost, and F is the feed flowrate. Finally, $C_{\mathrm{pump}} = \Sigma C_{\mathrm{pus}}\, Q_{\mathrm{s}}$, where C_{pump} is the pumping cost, C_{ps} is the unitary pumping cost and Q_{s} the flowrate in each stage s.

Due to the plant is not operating a fraction of time, the effective plant operational time EPOT is calculated as: EPOT = total time available (8760 h per year) — planned maintenance downtime (here assumed as 168 h) — estimated unplanned maintenance downtime (inherent unavailability, Eq. (20)). So, if we must produce Prod m^3/year, the net or real capacity (RProd m^3/year) must be:

$$R\mathrm{Prod} = \mathrm{Prod}/\mathrm{EPOT} = R\mathrm{Prod}/A_{\mathrm{sys}} \tag{18}$$

Where A_{sys} is the operative availability. It is calculated as $A_{\mathrm{sys}} = A_{\mathrm{sis}} * A_{\mathrm{sch}}$, where A_{sch} is the scheduled maintenance downtime (here assumed known), and A_{sis} is the inherent availability of the plant (unscheduled maintenance dowtime). RProd is the real capactiy of the plan when it is operative, or available (the base equipment are designed). Instead, Prod is the desired annual production, and here is a known data, a parameter of the problem. Note that always RProd will be greater than Prod, because we divide for a factor minor than one, in Eq. (18). So, the investment cost is increased according to the availability is decreased, to satisfy the desired annual production. Operative costs instead are the same, because fluxes increment in the same proportion as the operative time is shorten.

In reliability models, is implicitly assumed that during the time the system is unavailable, is repaired. So, the estimated corrective maintenance cost is:

$$C_{\mathrm{cmc}} = \Sigma\, m_{\mathrm{s}}\, C_{\mathrm{cms}}\, (1 - A_{\mathrm{s}})$$

Where C_{cms} is the annual corrective maintenance cost of equipment in stage s, m_{s} is the number of equipment in stage *s*, and A_{s} is the availability of stage *s*. The total corrective maintenance cost is achieved as the summation of all stage costs.

4. Reliability, maintainability and availability models

Availability of an equipment (i) (A_{i}) is the capacity of an equipment (or component or a process) to perform its design functions as the design specifications (given the "normal" environmental and operational conditions). By definition we have:

$$A_{\mathrm{i}} = \frac{\mathrm{MTTFi}}{\mathrm{MTTFi} + \mathrm{MTTRi}} \tag{19}$$

where MTTFi and MTTRi are the mean time to failure and mean time to reparation of each equipment respectively. In order to calculate both quantities we must adopt a reliability model for the system. Mathematically we can write by definition (see Henley and Kumamoto, 1981) $R(t) + F(t) = 1$. Where $F(t)$ is de unreliability function. In general, $R(t) = \exp\left(-\int_0^t r(t)\mathrm{d}t\right)$, $r(t)$ is called hazard function or hazard rate, and is a pdf of the failure rate. So, it is possible to derive the corresponding reliability and failure rate functions for different probability density functions.

Even it is known that failures rates are high in the initial period of operation, decreasing to a constant value at a "steady state operation", and also then grows in the final period of the life cycle, in general for continuous processes, at the conceptual design stage, it is convenient to assume a constant failure rate, to greatly simplify the problem. In fact, if we assume a constant hazard rate function $r(t) = \lambda$, we obtain the known exponential pdf for reliability $R(t)$. The advantage is a big simplification, because it is possible to achieve extremely simple relationships among the principal parameters describing the reliability and availability of the system. In fact, the MTTF can be obtained as follows, which for exponential distribution becomes time-independent

$$\mathrm{MTBF} = \int_0^\infty R(t)\mathrm{d}t = \int_0^\infty e^{-zt}\mathrm{d}t = \frac{1}{\lambda} \tag{20}$$

Maintainability $M(t)$ is a measure of the speed with which an equipment can be restored to operational status following a failure or removal from operation for servicing. Also can be defined as the probability that the equipment can be kept in an operational condition or restored to that condition within a given time when a defined maintenance schedule, and operational environment is given. Maintainability is often confused with maintenance, which is a series of specific actions taken to restore a machine to full operational status. These actions may include servicing, inspections, adjustment, lubrication, removal or replacement or repair in-place. Preventive maintenances are actions taken to retain the equipment at a given level of performance. Corrective maintenances are actions taken to restore the equipment to the operational status, after for example a failure. The same relationships used for reliability can be introduced for maintainability functions. If we assume a constant maintainability rate $\mu(t) = m$, now using the same definitions and procedures above described for reliability functions, we obtain:

$$\mathrm{MTTR} = 1/m \tag{21}$$

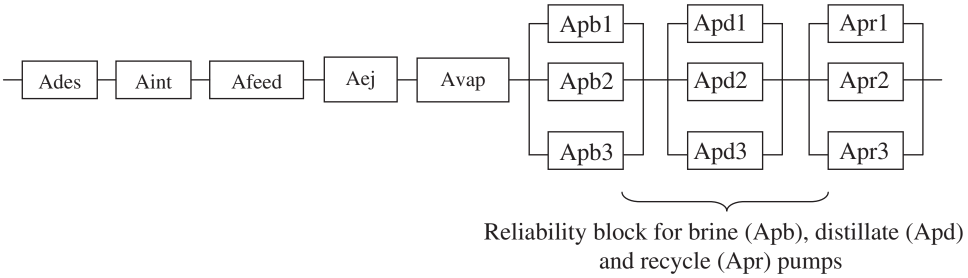

Fig. 2. Reliability block for the optimization problem.

The process or system availability (A_{sis}) must be calculated as a function of equipment availabilities (A_i), according to the equipment arrangement, which is captured by the system (process) availability model (considering the structure of the system: parallel, series, redundancy levels, etc). So, in a rigorous sense, we must introduce at this point, a general relationship between the process (Eqs. (1)–(16)) and the logic superstructure representing equipment allocation to calculate the system availability.

According to Fig. 2 our process can be represented by a series-parallel system combination. So, as all N-stage series systems, all components must be operating to ensure the process (system) operation. So, the system is unavailable (not operating, failed) when any one of the N stages (equipment or array of equipment) is not operating. For such (process) system, the availability (A_{p}) as a function of the availability of each stage (A_{s}) is:

$$A_{\mathrm{p}} = \prod_{\mathrm{s}=1}^{N} A_{\mathrm{s}} \tag{22}$$

For parallel arranges instead, the system is available if one of them is available. In this case, the system availability is calculated according to the following relationship:

$$A_{\mathrm{par}} = 1 - \prod_{\mathrm{js}=1}^{ms} (1 - A_{\mathrm{js}}) \tag{23}$$

j_{s} represents the availability of each equipment in parallel, and ms is the number of equipment in parallel in each stage. It is assumed that all the equipment in each stage are identical so Eq. (23) becomes $A_{\mathrm{par}} = 1 - (1 - A_{\mathrm{p}})^{ms}$. We introduce equipment redundancy by stand by arrangement. The above equations are valid assuming statistical independence among components. In warm stand by, it is assumed that the failure rate of stand by units (response to demand) is different to failure rates of the same equipment, but working as principal component. So, there is a statistical dependence among them, and the availability must be calculated in a different form that conventional parallel systems, Eq. (23). Here we use this equation to simplify calculations, and also because the error so introduced is very small.

5. Study case

In this section, a specific design problem considering a MSF-M process will be solved. Design parameters are given in Table 1. Costs data are given in Table 2. Availability data are given in Table 3. Specific data taken from literature are here adopted.

The following are all the restrictions we have used in the problem formulation: $A_{\min} \le A_{\mathrm{sys}}$.

In each stage, we permit at most a principal unit and two parallel pumps in stand by. So m_{s} is restricted by: $1 \le m_{\mathrm{s}} \le 3$. For stages in which stand by units are allocated, m_{s} is a variable of the problem. As each stand-by pump is identical, Eq. (23) is now $A_{\mathrm{pstby}} = 1 - (1 - A_{\mathrm{punit}})^{m_{\mathrm{s}}}$. Due to m_{s} is an integer exponent, a MINLP problem

Table 1
Parameters values and cost data for the optimization problem

Parameter	Value
Cp (heat capacity) [Kcal/Kg hr^{-1} K^{-1}]	1.3
U (heat transfer coefficient) [Kcal/m^2 hr^{-1} K^{-1}]	2200
Latent heat[Kcal/Kg]	570
Irreversibilities (BPE+NEA) [K]	2
Tmax (maximun temperature) [K]	393 K
Feed temperature[K]	298 K
Production (Prod)[tn/hr]	1000
F factor	20
DT (Tube diameter)[m]	0.030
CRF (Cap.Rec. Factor) [year^{-1}]	0.16
C_A (Heat transfer area cost) [\$/m^2]	50
CQ^{Des}(Hot utility cost) [\$/10^6 Btu]	0.252
Avap	0.99090
Aint	0.98138
Ades	0.97808
Aej	0.98900
Afeed	0.9788
Aop	0.9785
λ_{bri} year^{-1}	1.445
μ_{bri} year^{-1}	700
λ_{des} year^{-1}	1.0
μ_{des} year^{-1}	800
λ_{re} year^{-1}	1.234
μ_{re} year^{-1}	800

must be introduced adding binary variables through the following set of restrictions:

$$A_{pstby} = \text{apdes}_1 * x_1 + \text{apdes}_2 * x_2 + \text{apdes}_3 * x_3$$

$$\text{apdes}_1 = 1 - (1 - A_{punit})$$

$$\text{apdes}_2 = 1 - (1 - A_{punit})^2$$

$$\text{apdes}_3 = 1 - (1 - A_{punit})^3$$

$$x_1 = y_1,\ x_2 = y_2,\ x_3 = y_3,$$

$$y_1 + y_2 + y_3 \geq 1$$

Where A_{punit} is the availability of each pump, calculated using Eq. (19) and x_i (real and continuous variables) and y_i (binary variables) are auxiliary variables to decide the number of pumps to be allocated (m_s). This set of equations must be introduced for each stand by stage in the reliability block (Fig. 2). In this case, distillate, brine, recirculation and feed pumps.

Table 2
Optimal values obtained by minimizing TAC (process availability free)

Variable	Values
Q^{Des} [Gcal/hr]	80.079
Nt	2494
NS	33
Vel [m/s]	1.305
Δtc [K]	8.207
Vbri [m/s]	3.000
WR [tn/hr]	5.708
Δtf [K]	2.585
Δte [K]	10.793
At [m^2]	38658.34
As[m^2]	33129.10
L [m]	56. 743
Asys	0.897
Asis	0.916
Asch	0.978
Recicle pumps	1
Brine Pumps	1
Distillate pumps	1
TAC [with availability, 10^6 \$/year]	1.212
TAC [without availability, 10^6 \$/year]	1.3267
H[m]	2.742
Htu [m]	1.742
Rpro[tn/hr]	1113. 5
Prod[tn/hr]	1000
% utility cost	49.6
% area cost	45.2
% pumping cost	3.5
% Corrective maintenance cost	1.7

6. Results and discussion

In this section we analyze different results for a given example. In Table 1 we show the main design parameters of the process and availability data for each equipment (stage) of Fig. 2 [17–18].

Table 3
Optimal values obtained by maximizing the process availability (TAC free)

Variable	Maximizing Asys	Minimizing TAC
Asys	0,901	0,8987
Asis	0,921	0,919
TAC [\$/year]	$1{,}4821 \times 10^6$	$1{,}3267 \times 10^6$
Asch	0,978	0,978
Brine pumps	3	1
Distillate pumps	3	1
Recicle pumps	3	1
Rprod	1100,0	1113,5

In first place, we solve the optimization problem minimizing the investment cost with the process availability free. It is important to remember that if A_{sis} decrease, the real capacity of the process must increase in order to satisfy the annual demand. There is a tradeoff between the increase of investment in redundancy (more standby equipments) to improve the availability of the plant and the extra capital cost to increment equipment capacity.

The problem was solved using GAMS, the execution time was 0.031 sec (CPU time). In Table 2 the principal results are shown.

If we now solve the problem Max A_{sys}, using the same above model, the maximum availability for the process is achieved, increasing of course the capital investment. In Table 3 are given same differences between both solutions. It is clear that with this parameters, the minimum cost is $1.3267 \cdot 10^6$, but the availability is 0.897. On the other hand, the maximum availability is 0.919, but the cost is $1.4821 \cdot 10^6$.

7. Conclusions

In this paper a methodology for a MSF-M process optimization taken into account process availability was introduced. The process optimal operative conditions, and the optimal equipment allocation in stand-by units is achieved, given a process model and cost data. It is shown that maintenance and reliability play an important role in process operability and productivity.

The problem is modeled as a MINLP problem. The stand-by units allocated in each pumping stage introduce integer variables as exponents. The problem is transformed in a MINLP program introducing binary variables and suitable restrictions which captures the more important variables of the process, and it relationships. Both, the process and the availability models are integrated and solved easily according to the proposed strategy. The model is flexible and robust. We solved the minimization of total cost, or the maximization of the process availability. Also, in both problems we can introduce the natural restrictions on a maximum available capital, or minimal availability, achieving solutions between these extreme operating conditions.

Several points remains for future works. On one hand, a better, more rigorous process model must be used; not only for MSF-M, but for conventional MSF processes also. On the other hand, better models for availability must be introduced, in particular, the relationships among reparation times and corrective maintenance costs. In addition, preventive maintenance can be introduced. Also, more rigorous equations for calculation of stand-by equipment availability must be contemplated.

Of course, if different process structures (a superstructure) is considered for the process design (synthesis), each option explored during the search must interact with the availability block in order to calculate the total process cost. So, a link among both models must be introduced. This also will surely introduce a much more complicated MINLP program to be solved. However, this first step proved to be adequate both, in model representation and programming efficiency to solve the problem. In future works, all of these open problems will be incorporated to the existing software step by step.

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References

[1] C. Cordier, M. Fayot, A. Leroy and A. Petit, Integration of process simulations in availability studies, Rel. Eng. Sys. Safety, 55 (1997) 105–116.

[2] J.M. Douglas, Conceptual design of chemical processes, New York, McGraw Hill, 1988.

[3] J.K. Grievink, K. Smit, R. Dekker and C.F.H. Van Rijn, Managing reliability and maintenance in the process industry, In: Conference on Foundation of Computer Aided Operations, FOCAP-O, 1993, pp. 133–157.

[4] M. Hajeeh and D. Chaudhuri, Reliability and Availability assessment of reverse osmosis, Desalination, 130 (2000) 185–192.

[5] E.J. Henley and S.L. Gandhi, Process reliability analysis, Am. Inst. Chem. Eng. J., 21 (1975) 677–686.

[6] N. Ishii, T. Fuchino and M. Muraki, Life cycle orientated process synthesis at conceptual planning phase, Comp. Chem. Eng., 21 (Suppl.) (1997) S953–S958.

[7] W. Kuo, V.R. Prasad, F.A. Tillman and C. Hwang, Optimal reliability design, Cambridge University Press, 2001.

[8] L.A. Painton and J. Campbell, Genetic algorithms in optimization of system reliability, IEEE Trans. Reliab., 44 (2) (1995) 172–178.

[9] J.S. Tan and M.A. Kramer, A general framework for preventive maintenance optimization in chemical process operations, Comp. Chem. Eng., 21 (12) (1997) 1451–1469.

[10] G. Thangamani, T.T. Narendran and R. Subramanian, Assessment of availability of a fluid catalytic cracking unit through simulation, Reliab. Eng. Sys. Safety, 47 (1995) 207–220.

[11] T.V. Thomaidis and E.N. Pistikopoulos, Optimal design of flexible & reliable process systems, IEEE Trans. Reliab., 44 (2) (1995) 243–250.

[12] C.G. Vassiliadis and E.N. Pistikopoulos, On the interactions of chemical-process design under uncertainty and maintenance optimization, Proceedings Annual Reliability and Maintainability Symposium, 1998, pp. 302–307.

[13] C.G. Vassiliadis and E.N. Pistikopoulos, Chemical-process design and maintenance optimization under uncertainty: a simultaneous approach, In: Proceedings Annual Reliability and Maintainability Symposium, 1999, pp. 78–83.

[14] H.D. Goel J., Grievink, P.M. Herder and M.P.C. Weijnen, Reliability Engineering and System Safety, 78:3 (December 2002) 247–258.

[15] N. Scenna, Simulación y Síntesis de Procesos Químicos. Su Aplicación a Procesos Térmicos de Desalación, PhD Thesis, Universidad Nacional del Litoral, Santa Fe, Argentina, 1987.

[16] H. El-Dessouky, H. Shaban and H. Al-Ramadan, “Steady-state analysis of multi-stage flash desalination process,” Desalination J., 103 (1995) 271.

[17] OREDA, Offshore reliability data handbook, Norway: OREDA Participants, 1984.

[18] C.F.H. Van Rijn, A systems engineering approach to reliability, availability, and maintenance, Park City, UT, 1987.