

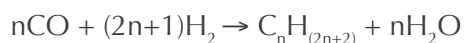


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INTENSIFIED FISCHER-TROPSCH SYNTHESIS BY A MONOLITHIC REACTOR

A Fischer Tropsch monolithic reactor is simulated by a heterogeneous dispersed steady-state model; material transports are described by a film model based on Maxwell-Stefan equations. This model shows that a Fischer-Tropsch monolithic reactor can be surprisingly sensitive to some parameters; in particular, with a productivity of industrial practical interest, reactor can exhibit state multiplicity. Just very large unrealistic metallic matrix thermal conductivity can assure state unicity and stability; otherwise reactor control could be a difficult task. In many papers where a monolithic reactor is proposed for industrial developments of the Fischer-Tropsch process, no stability analysis is performed, though this is a crucial problem for an intensified process. However multi-tubular reactors loaded with washcoated structured catalysts seem to be stable; instead, if channels are filled with catalytic particles, our model evinces hot-spot in axial temperature when the reactor get productivity values of industrial practical interest.

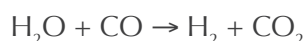
The Fischer-Tropsch process is a catalyzed chemical reaction in which a mixture of carbon monoxide and hydrogen (syngas) is converted into liquid hydrocarbons of various forms. The resulting overall reaction can be presented as follows:



being:



An important reaction to take into account is the water gas shift reaction:



There is two different process: a low temperature process (LTFT), with temperatures ranging between 200-240 °C and a high temperature process (HTFT), with temperatures ranging between 300-350 °C. The

HTFT uses an iron catalyst, and the LTFT either an iron or a cobalt catalyst. Other catalysts are not used for their high price.

In general the product distribution of hydrocarbons formed during the Fischer-Tropsch process follows some sort of exponential function, with the probability of chain growth as an important factor. There are two main methodologies to predict product distribution: a kinetic approach and a thermodynamic approach. The model called "Anderson-Schultz-Flory" (ASF) distribution model is the best known thermodynamic model. It was developed in the early years of FT-synthesis. ASF model can be expressed as:

$$W_n/n = (1-\alpha)^2 \alpha^{n-1}$$

where W_n is the weight fraction of hydrocarbon molecules containing n carbon atoms. α is the chain growth probability or the probability that a molecule will continue reacting to form a longer chain. Increasing α increases the formation of long-chained



hydrocarbons. In general, α depends on the catalyst and the specific process conditions.

Kinetic models are based on detailed FTS mechanism explaining the synthesis of each product through the evolution of reaction intermediates and adsorbed species [1].

Process conditions and catalyst are usually chosen to promote higher order reactions and to minimize methane formation. Most of the paraffins produced are straight-chained, although some branched paraffins are also formed. In addition, several competing reactions form olefins, as well as alcohols and other oxygenated hydrocarbons.

All reactions taking place in the reactor are highly exothermic, and to avoid an increase in temperature, which results in lighter hydrocarbons and carbon deposition on the catalyst, it is important to have sufficient cooling, to secure stable reaction conditions. For industrial FT reactors heat removal for temperature control is the most important feature to obtain optimum product selectivity, long catalyst life and stable regime.

Catalyst pellet-loaded fixed bed and catalyst slurry are the main reactors commercially used; sometime fixed fluidized bed and circulating fluidized bed reactors are used too. In the last decade several alternative reactor technologies have been analyzed and developed. In particular there has been growing attention for honeycomb monolith reactors consisting of long small parallel channel [2-8].

Advantages and drawback of each industrial reactor type have been largely studied [9, 10].

In fixed-bed reactors process is carried out at high linear velocity to control heat removal rate. Large catalyst particles are used to reduce pressure drop: then reaction is controlled by intra-particle diffusion; often, to reduce pore-diffusion effects, pellets with a thin catalyst outer layer are used. In any case fixed-bed reactors suffer of low productivity, high methane formation, low α values and short catalyst life.

Slurry reactors provide several advantages: intra-particle diffusion problem is solved by using small catalyst particles, productivity is higher, design is simple and catalyst life is longer. However separation of liquid products from the catalyst, inter-phase gas-liquid mass transfer and back-mixing are great problems of these reactors.

Fluidized bed and slurry reactors advantages and drawbacks are similar, but fluidized bed reactor design and control are more complex.

Catalytic monolith reactors are widely applied in environmental catalysis [11-13] and, in particular, in automotive catalytic converters to reduce vehicles emissions [14-16]. By the other side the use of catalytic monolith reactors in chemical processes has been quite limited till today [17-23].

Application of monolith reactors for three-phase reactors and in particular for FT synthesis is very promising due to their several features [24, 6, 25, 26]:

- low pressure drop;
- uniform liquid distribution;
- smaller reactor's size (process intensification);
- high heat removal rate;
- high olefin to paraffin ratios.

The main disadvantage is, however, the lack of practical experience with monolithic reactors. Moreover, a monolithic reactor is a very complex system from the modeling view point [27, 28]. Experimental and theoretical efforts have to be pursued to further investigate behavior of these reactors. In particular, mathematical models can be a very useful tool to gain insight for the analysis and the rational scale-up and design of the monolithic reactors. So we have developed a detailed and general steady state mathematical model of a three-phases FT monolithic reactor based on honeycomb metallic supports coated by catalyst or filled with little catalyst pellets. The model has been implemented in an equation oriented simulator [29] and its reliability has been tested on a LTFT example reported in a fairly recent patent [30]. Our model results agree quite well with the values reported in that example, but they showed that the stability domain of such reactors is so small that a further dynamic analysis should be recommended, for proper design and management.

Monolithic reactor configuration [10, 25, 28, 30]

A catalytic monolith reactor is a multi-tubular cylindrical reactor; a refrigeration fluid is flowing inside the reactor, externally to the tubes. Each internal cylindrical tube contains an honeycomb monolith substrate having many parallel channels separated by thin walls. Each longitudinal channels has a cross section in the form of a square, rectangular, circular, or a more complex form, with 0.5 to 5 mm length. In the practical applications a standard honeycomb monolith substrate has 200-500 cells per square inch (cpsi); for a standard monolith tube, generally, diameter ranges from 2 to 4 cm, length from 10 to 50

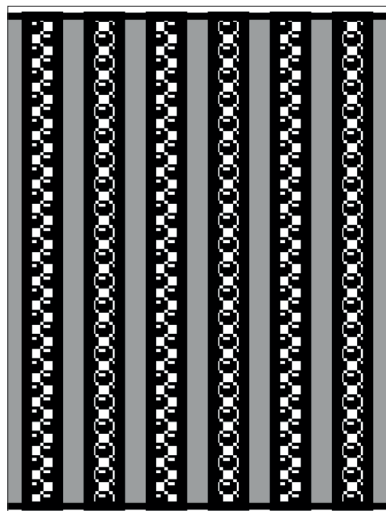


Fig. 1 - Reactor scale interpretative model: view

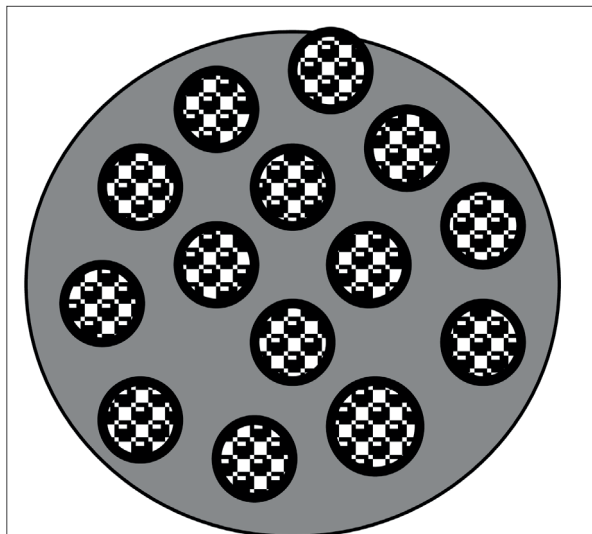


Fig. 2 - Reactor scale interpretative model: plant

cm. Reactor height and diameter range respectively from 1 to 10 m and from 1 to 5 m. The internal tube-monolith tolerance must be very small because it is relevant to ensure good heat transfer U to the cooling fluid. Channel shape, void fraction and material properties of the monolith ensure a significantly higher thermal conductivity compared with that of a packed bed. Catalyst is coated onto the surface of the metallic matrix or fills micro-channel as little pellets or structured bodies. In both cases, thank to short diffusion lengths, selectivity on heavy hydrocarbons is achieved [10, 31, 32]. Reactors with pellets or structured bodies inside have an higher volumetric density of the active phase than that of the reactors with “washcoated”. Then reactors with pellets or structured bodies present very higher productivity as long as they have an efficient system for the removal of the reaction heat. For this purpose a material with high thermal conductivity (like copper or aluminum), has to be used for tubes and monolith substrate. So radial heat transport can be more favorable and risk of thermal runaway and catalyst deactivation could be reduced [21].

Model structure

Dimensions of the micro-scale (cells) and of the macro-scale (monolith tube) differ by several orders of magnitude. By a mathematical continuous model we can consider mass conservation equations in a representative elementary volume (REV) V , where V is a volume that is large with respect to the cell

dimension, but small compared to the dimensions of the monolith tube. In V we consider each fluid phase homogeneous and embedded within the honeycomb monolith. In order to use the infinitesimal analysis rules, generally V is dealt as an infinitesimal and indicated as δV , by supposing that all properties are continuous and derivable functions of V . By this way, on the basis of the divergence

theorem and some simple analysis rules, any conservation equation in the V space become a differential equation in the dV space. This approach uses a very punctual internal physical detail (“multiple gradient model”) with a scale smaller than original REV scale. Chosen a suitable number of points (“grid”), space derivatives are replaced by difference approximations on this point. The resulting nonlinear difference equations are solved by either fully implicit or semi-implicit techniques. Grid block has to be chosen to satisfy both numerical stability and internal physical detail. Numerical methods usually applied in space and time discretization are orthogonal collocation, orthogonal collocation on finite elements, Galerkin finite elements. All these numerical methods are normally unbounded, which means that unphysical oscillations can appear in the computed solution, with numerical dispersion of temperature and concentration profiles. A number of developed discretization methods claims to be able to minimize these effects [33-37]. We prefer to use lumped equations directly, without converting them into differential equations and choosing the grid (each V size) on the basis of geometrical and structural details that are effectively known. By this way mass conservation equation for a component i becomes an algebraic equation.

Model formulation

Reactor is described by an interpretative model (Fig. 1, 2) based on different description levels for its four

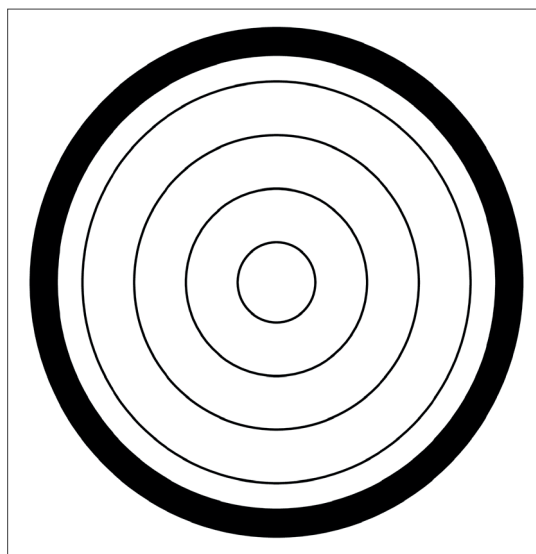


Fig. 3 - Reactor scale interpretative model: macro-channels

fundamental macroscopic elements ((full reactor, monolith tube, monolith macro-channels and “cells”).

At reactor scale, model is formulated as N_{MON} parallel tubes all interacting with a refrigerant fluid.

Thank to tubular monolith azimuthal symmetry, in each tube, behavior of each monolith channel along a radius is invariant with the radius we consider. Then, at monolith tube scale, model can be formulated as N_{CHAN} concentric annular macro-channels (Fig. 3), where each macro-channel i , between radius r_{i-1} and radius r_i , contains:

$$NF_{\text{CHAN},i}^{\circ} = NF_{\text{CHAN}} \{(r_i/r_0)^2 - [(r_{i-1})/r_0]^2\}$$

physical channel, being NF_{CHAN} the total channel number contained in each tubular monolith and r_0 the tubular monolith radius.

Global flow F of the gas and liquid streams fed to the monolithic reactor is partitioned among N_{MON} equal streams to each monolith tube, with F/N_{MON} rate; each of these streams are then partitioned among N_{CHAN} streams to each concentric annular macro-channel; each stream rate is proportional to the corresponding annular section area.

By our approach [38] axial mixing effect in each annular macro-channel is described by partitioning the channel in a suitable number of “cells” (grid elements) in cascade. NP grid elements number is chosen on the basis of longitudinal Peclet values and geometrical and structural details [29]. To distinguish between

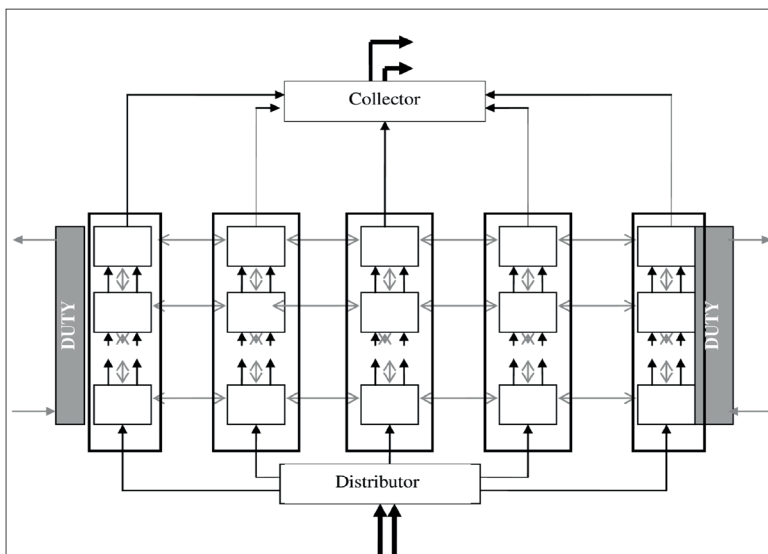


Fig. 4 - Monolithic reactor scale interpretative model (5 macro-channels)

on the two phases (gas and liquid) axial mixing, we introduce, for each phase, a different back flow. We neglect radial gradients and consider bubble composition in each “cell” as uniform.

The monolithic reactor interpretative model corresponds to the one of a single monolith tube (Fig. 4) fed by a flow with rate $F_{\text{MON}} = F/N_{\text{MON}}$.

By a “distributor” we divide this flow among N_{CHAN} flows to the macro-channels, where the i^{th} macro-channel fed rate is

$$F_{\text{CHAN},i}^r = F/N_{\text{mon}} \{(r_i/r_0)^2 - [(r_{i-1})/r_0]^2\}$$

By a “collector” we collect all the N_{CHAN} streams $q_{\text{CHAN},i}^r$ coming out in one monolith tube stream $q_{\text{MON}} = \sum_i q_{\text{CHAN},i}^r$. Flow rate from the reactor is equal to $q = N_{\text{MON}} q_{\text{MON}}$.

At grid element (“cells”) scale we introduce the following main hypothesis:

- mass transport between fluid phases and liquid-solid phases is described by a transfer-based model, with reference to a film model, where molar fluxes are calculated by the Maxwell-Stefan equations [29];
- vapor phase doesn't exchange with solid phase;
- active catalytic sites are present just on the catalyst surface.

On the basis of these hypothesis the grid single element model depends on the following $6N_{\text{cp}} + 9 + N_{\text{RB}}$ variables according to Tab. 1, being:

Name	Description	Number
$x=x_i, i=1, N_{cp}$	Liquid molar fractions	N_{cp}
$y=y_i, i=1, N_{cp}$	Vapour molar fractions	N_{cp}
T	Temperature [K]	1
V	Vapour molar rate [Kmol/h]	1
L	Liquid molar rate [Kmol/h]	1
$x_s=x_{si}, i=1, N_{cp}$	Surface catalyst molar fractions	N_{cp}
P	Pressure [atm]	1
T_{UY}	Duty temperature [K]	1
E_B	Enhancement factor of equilibrium reactions [Kmol/h]	NRB
$xy^l=xy^l_i, i=1, N_{cp}$	Liquid molar fractions on LV interface, liquid side	N_{cp}
$yx^l=yx^l_i, i=1, N_{cp}$	Vapour molar fractions on VL interface, vapour side	N_{cp}
Φ^{tLV}	Total mass transfer rate on LV interface, liquid side [Kmol/h]	1
Φ^{tLS}	Total mass transfer rate on LS interface, liquid side [Kmol/h]	1
Φ^{tVL}	Total mass transfer rate on LV interface, vapour side [Kmol/h]	1
$xxs^l=xxs^l_i, i=1, N_{cp}$	Liquid molar fractions on LS interface, solid side	N_{cp}
T^s	Solid phase temperature [K]	1

Tab. 1 - Model variables

N_{cp} : chemical component number

N_{RB} : number of enhancement factor of equilibrium reactions.

The grid single element model is implemented by the mathematical formulation of $6N_{cp}+9+NRB$ statements (Tab. 2), where multicomponent Maxwell-Stefan mass transfer flux values Φ^{LV}_i , Φ^{VL}_i , Φ^{LS}_i can be obtained by solving Maxwell-Stefan equations, with reference to a film model [51].

In particular, in equation 18 (Tab. 2), radial heat transfer rate at monolith radius r is estimated by

$$2\pi k_{e,r}(1-\varepsilon)\Delta z[(T^s_r + \Delta r - T^s_r)/\ln r/(r+\Delta r)]$$

and axial heat transfer rate at monolith height z is estimated by

$$S(1-\varepsilon)k_{e,z}[(T^s_z - T^s_z + \Delta z)/\Delta z]$$

being

$k_{e,r}$: matrix effective radial thermal conductivity

ε : matrix void fraction

Δz : "cell" axial dimension

Δr : "cell" radial dimension

r : radial coordinate

T^s : solid phase absolute temperature

$k_{e,z}$: matrix effective axial thermal conductivity.

Formulation and numerical solution behavior of the monolith model benefit from equation oriented architecture of the simulator [38-43].

Model application and sensitivity evidence

A multi-tubular monolithic reactor is of industrial practical interest for the Fischer-Tropsch reaction just if its productivity is very higher than the traditional slurry or fixed bed reactors, so, with the same production, reactor has very smaller dimension [30]. This condition can justify the very high investment for this quite new technology.

We tested our model on an externally cooled monolith mono-tubular cylindrical reactor made of aluminium, where reactor design and operational conditions are reported in Tab. 3 and are quite similar to those described in an international patent [30]. As suggested in this patent, in order to have an appreciable productivity, channels are filled with catalytic particles packed with cobalt supported on alumina having an external diameter of 500 μm [30]. Catalyst body dimension has to be quite small to reduce intra-particle mass transfer resistance, responsible for a low activity and selectivity, but not too small to avoid large pressure drop. "Washcoated" monolithic catalysts (channels coated with a layer of catalyst) are less promising because require larger



Description	Number
1) Component material balances in vapour phase	Ncp
2) Component material balances in all phases	Ncp
3) Sum equations for x-y	1
4) Energy balance for fluid phases	1
5) Global material balance in all phases	1
6) Sum equations for xy^l	1
7) No solid intra-phase resistance: $x_{si} = x_{sl}$	Ncp
8) Energy balance on Duty	1
9) Momentum balance	1
10) Chemical equilibrium relations	N_{RB}
11) LV Inter-phase Equilibrium relations	Ncp
12) Continuity on LV inter-phase fluxes: $\Phi^{LV}_i = \Phi^{VL}_i$, $i(\neq \Lambda) = 1, Ncp$	Ncp-1
13) Sum equations for yx^l	1
14) Continuity on LS inter-phase total: $\Phi^{LS} = R_t$	1
15) Continuity on LV inter-phase total flux: $\Phi^{LV} = \Phi^{VL}$	1
16) Continuity on LS inter-phase fluxes: $\Phi^{LS}_i = R_i$, $i(\neq \Lambda) = 1, Ncp$	Ncp-1
17) Sum equations for xx^l	1
18) Energy balance for solid phase	1

Tab. 2 - Mathematical model statements

volumes in order to have an appreciable production. Catalyst thermal conductivity is neglected. Effective matrix axial thermal conductivity $k_{e,z}$ is supposed equal to metal thermal conductivity k corrected by matrix void fraction; effective matrix radial thermal conductivity $k_{e,r}$ is calculated by a predictive equation [26, 44].

We have gotten chemical-physical and kinetic equations from literature correlations, as reported in Tab. 4.

Kinetic model based on detailed mechanism explaining the synthesis of each product through the evolution of reaction intermediates and adsorbed species [45] are very useful for the analysis and development of the process where a correct prediction of the experimental product distribution is fundamental. For a thermal analysis we prefer

per reactor's volume unit allow to reach a total specific productivity of about 290 kg/h/m³ similar to the value reported in the cited patent [30], very higher

to use a simplified model with hydrocarbons and alcohols distribution determined by the ASF model.

The reactor model provides the following results in good agreement with the reference example (see Tab. 3).

Thanks to the pellet catalyst dimension less than 1 mm we obtain an acceptable pressure drop, less than 0.6 atm/m. So, being the removal of reaction heat efficient, thanks to the conductive monolithic structure, and a very elevated values assigned to the parameter U and to the thermal conductivity k , the quantities of catalyst

	Our simulation	Patent [30]
Void fraction	0.57	0.57
Pellets external diameter	0.5 mm	0.5 mm
Catalyst volume	550 cm ³	?
Monolith tube length	3 m	1 m
Monolith tube diameter	2.54 cm	2.54 cm
Channel diameter	0.5 cm	0.5 cm
Metal thermal conductivity k	175 kcal/m/h/K	100-400 W/m/h
Heat transfer to the cooling fluid U	600 kcal/h/m ² /K	?
Pressure	20 bar	20 bar
H ₂ /CO feed molar ratio	2.1	2.1
Evaporating water temperature	215 °C	228 °C
Feed rate	3.2 Nm ³ /h	?
Specific feed rate	3050 Ncm ³ /h/g cat	5000 Ncm ³ /h/g cat
Feed temperature	200 °C	228 °C
Pressure drop	0.6 atm/m	0.7 atm/m
Productivity	285 kg/h/m ³	264 kg/h/m ³

Tab. 3 - Reactor design, operational conditions and performance

Fugacity coefficients	Redlich-Kwong-Soave [51] with modified Huron-Vidal mixing rules [52]
Standard fugacity	Virial equation of state [53]
Vapour pressure	Riedel equation [54]
Liquid and vapour viscosity	Dean and Stiel correlation [55]
Surface tension	Macleod-Sugden correlation [56, 57]
Vapour binary diffusion coefficients	Chapman-Eskog theory [58]
Liquid binary diffusion coefficients	Scheibel-Vignes correlations [59, 60]
Enthalpy	Redlich-Kwong-Soave [51] with modified Huron-Vidal mixing rules [52]
Vaporization heat	Watson method [61]
Kinetic equation for the conversion of CO and H ₂	Yates model [62]

Tab. 4 - Chemical-physical and kinetic correlations

than the traditional slurry or fixed bed reactors. At design conditions, radial temperature profile is rather flat, thank to elevated thermal conductivity; high hot spots are avoided in longitudinal temperature profiles as show in Fig. 5 where the abscissa values represent the grid number, being 1 the inlet cell. In Fig. 6 are reported the conversion rate profiles of H₂. H₂ conversion is higher at the reactor wall than

along the axis, thank to the increasing selectivity to heavier hydrocarbons with temperature drop. However we must pay some attention to the H₂ conversion steep ascent.

A fundamental condition for a feasible industrial plant scale development concerns the reactor stability and then its “controllability”. In fact, since a monolithic Fischer-Tropsch reactor is a rather new solution for Fischer-Tropsch synthesis [50], not very large knowledge exists on the operational behavior of the system. Therefore a controllability analysis has priority in the development of the reactor system, especially if they should operate on board ships, as suggest in the cited patent [30].

Monolithic reactors with metallic matrix where highly exothermic reactions take place are in effect “complex systems”, i.e. systems that, for their high non linearity, may have unexpected and not trivial model behavior [46, 47].

Rather recent practical experience with monolithic reactors and their degree of complexity require great experimental and theoretical efforts to investigate their behavior.

In traditional fixed bed reactors where longitudinal and radial dispersion are present, in particular when fluids flow upstream, multiple steady states can exist, even if, in general, design parameter values can produce a stable steady-state condition.

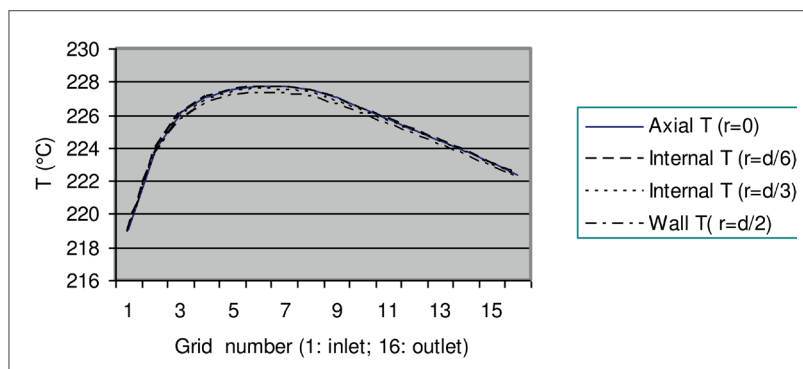


Fig. 5 - T longitudinal profile at design condition (4 radial positions)

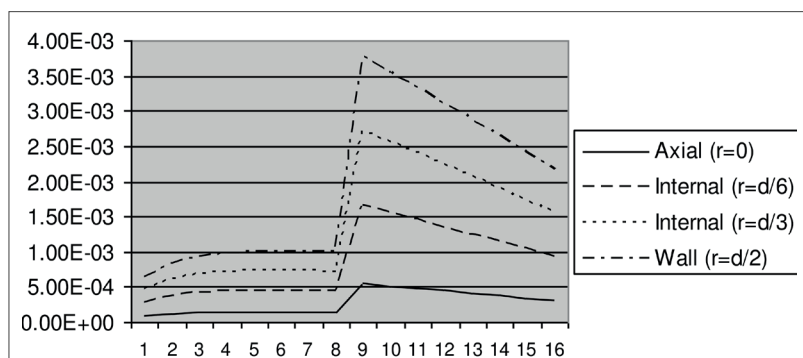


Fig. 6 - Longitudinal profile of hydrogen conversion rate at design condition (4 radial positions)

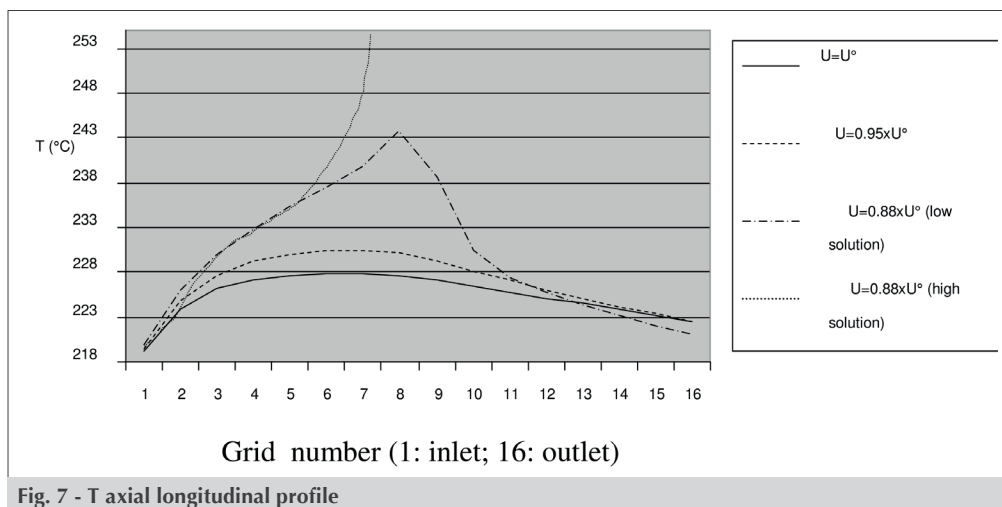


Fig. 7 - T axial longitudinal profile

Honeycomb monolithic structure with small packed or structured bodies of catalytic material assures that in the channels there is a stable laminar flow with flat velocity profile. So, material dispersion is negligible and we can assume a plug flow inside the channels. Moreover, being the matrix thermal conductivity very high, radial temperature gradients are quite absent (Fig. 5), except very near the walls. So reactor behavior, except for the effects of longitudinal thermal conduction, can be described with a plug flow model [10]. It is easy to show that in plug flow reactors only unique steady state profiles are possible, but, with particular system parameter values, these profiles can be surprisingly sensitive; on the contrary, the cited patent [30] seems to avoid this situation.

However the problem requires further analysis. In fact Fischer-Tropsch synthesis is carried out in a three-phase system with quite complex kinetics; here matrix heat conductivity could allow a heat back-flow, with a true heat diffusion effect, from the hot reactor section to the feed section, where, owing to the H_2 high concentration and high diffusivity, the reaction velocity can greatly increase. So a new "attractor" could be present for monolithic dynamic system [39], corresponding, in a steady state condition, to an instability [48, 49] with a very difficult reactor control. Fortunately, even if the high productivity and the unavoidable heat transfer resistance on the monolith external surface could give rise to a parametric sensitivity to a small perturbations, the same thermal metallic conductivity, for suitable values, should contrast this situation by a heat axial forward-flow and a large heat radial flow.

To verify previous theoretical analysis, we have slightly modified two parameters with respect to their

design values:

- wall heat transfer coefficient U , i.e. heat exchange coefficient in the clearance between the monolith external wall and the internal monolith tube wall ("gap" [26]);

- matrix metal thermal conductivity k .

If we reduce design value U° (600 kcal/h/m²/K) of the "gap" U by 5% and by 12% (both values

below the design precision) we obtain the respective changes in axial longitudinal temperature profiles reported in Fig. 7.

These preliminary results show that a monolithic reactor performing the Fischer-Tropsch synthesis is sensitive with respect to the external heat-exchange coefficient U , when the reactor is designed to have a productivity of industrial practical interest with respect to the traditional reactors. Evolution from low to high solution, for $U=0.88 \times U^\circ$, cannot be analyzed by a steady state approach.

Just with very large, unrealistic, matrix metal thermal conductivity values $k=2k^\circ$ (being $k^\circ=175$ kcal/m/h/K aluminium thermal conductivity) sensitivity disappears (Fig. 8).

Very large (unrealistic) heat transfer areas* and/or conductivity are necessary to guarantee state unicity and stability. State multiplicity disappears in this case, but oscillations can occur [50].

Alternatively, a very efficient control structure have to be designed to stabilize the unstable operating points. This goal requires detailed dynamic models.

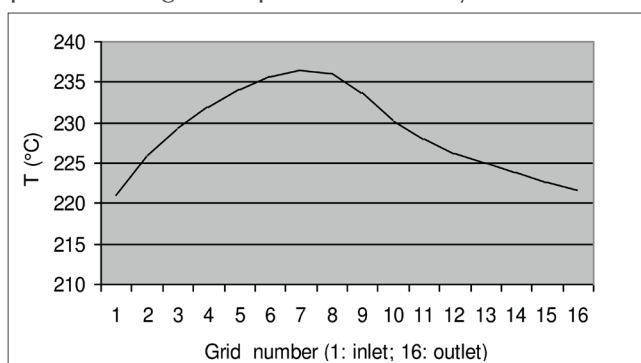


Fig. 8 - T axial longitudinal profile $U = 0.88 \times U^\circ$, $k = 2 \times k^\circ$

Conclusions

At first glance, there could appear to be little reason to include time dependence in a model when only steady states are of interest; so we have developed a detailed and general steady state mathematical model of a three-phases FT monolithic reactor implemented in an equation oriented simulator.

By this model we have simulated behavior of a pilot scale reactor like that described in a patent [30], obtaining similar results. But, on the basis of this analysis, it seems that high productivity FT monolithic reactors could work in a sensitive regime that should require a very efficient control system. So both our simulation results as other works [50] suggest that should be essential to develop not only steady state mathematical models, but dynamic models too, to perform rational and reliable scale-up, design and management of a high productivity FT monolithic reactor. In our opinion, until now this feature has been a little underestimated.

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Sintesi di Fischer-Tropsch intensificata da un reattore monolitico

Viene simulato, mediante un modello stazionario eterogeneo con dispersione assiale, un reattore pilota monolitico per la sintesi Fischer; il trasporto di materia viene descritto mediante un modello a film basato sulle equazioni di Maxwell-Stefan. Il modello mostra che un reattore monolitico per la sintesi Fischer-Tropsch può risultare notevolmente sensibile rispetto ad alcuni parametri di conduzione e/o di progetto; in particolare, per una produttività di interesse industriale, il reattore può presentare più stati stazionari di funzionamento. Solo un valore molto elevato, in pratica non realizzabile, della conduttività termica della matrice metallica può assicurare unicità di funzionamento e stabilità; altrimenti il reattore potrebbe presentare difficoltà di controllo. In molti lavori dove vengono proposti reattori monolitici per la realizzazione su scala industriale del processo Fischer-Tropsch viene trascurata l'analisi di stabilità degli stessi benché questo sia un problema fondamentale per un "intensified process". In base al nostro modello, un reattore multi-tubolare caricato con catalizzatore strutturato washcoated risulta stabile; tuttavia, quando i canali sono invece riempiti, allo scopo di ottenere una produttività di interesse industriale, con catalizzatore in pellet, il modello evidenzia un pericoloso hot-spot nel profilo assiale di temperatura.