Kinetics of long chain n-paraffin dehydrogenation over a commercial Pt-Sn-K-Mg/γ-Al₂O₃ catalyst


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Kinetics of long chain \( n \)-paraffin dehydrogenation over a commercial Pt-Sn-K-Mg/\( \gamma \)-Al\(_2\)O\(_3\) catalyst: Model studies using \( n \)-dodecane

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\( ^b \) Faculty of Science and Technology, University of Twente, 7500 AE, Enschede, The Netherlands
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Long chain paraffins
\( n \)-dodecane
Kinetic studies
Pt-Sn/Al\(_2\)O\(_3\)
Olefins

**A B S T R A C T**

A kinetic modeling study on long chain \( n \)-paraffin dehydrogenation using a commercial Pt-Sn-K-Mg/\( \gamma \)-Al\(_2\)O\(_3\) catalyst was carried out in a continuous flow set-up using \( n \)-dodecane as a model component at various temperatures (450–470 °C), pressures (0.17–0.30 MPa), \( \mathrm{H}_2 \)/paraffin mole ratios (3:1–6:1) and space times (0.22–1.57 g h mol\(^{-1} \)). The commercial catalyst was characterized by XRD, BET, MIP, SEM and CO chemisorption. An empirical exponential equation was found to predict the mono- and di-olefin yields very well. In addition, 6 mechanistic models based on the LHMW mechanism were derived and tested by non-linear least squares fitting of the experimental data. The model which assumes that surface reactions and particularly the dehydrogenation of the metal-alkyl chain to the adsorbed mono-olefin and di-olefin as the rate determining steps was found to give the best fit with the experimental data. In addition, activation energies and adsorption enthalpies for each elementary reaction were obtained. The kinetic testing and modeling have shown that the high mono-olefins selectivity for long chain paraffin dehydrogenation can be obtained by operating at low space time (when \( P \) and \( T \) and \( m \) are same), high pressure (when \( \tau \), \( T \) and \( m \) are same) and high \( \mathrm{H}_2 \)/paraffin ratio (when \( \tau \), \( P \) and \( T \) are same), as well as low reaction temperature (when \( \tau \), \( P \) and \( m \) are same) but with little effect.

1. Introduction

Dehydrogenation of long chain, kerosene range, \( n \)-paraffins to mono-olefins over a Pt-Sn/\( \gamma \)-Al\(_2\)O\(_3\) based catalyst operated at high temperatures (475–490 °C) and low pressures (0.1 – 0.25 MPa) is an important step in the production of linear alkylbenzene sulfonates (LAS), with widespread applications as biodegradable detergents. The commercial feedstock is a mixture of \( n \)-alkanes (generally \( n \)-C\(_{10}\)-C\(_{13}\)), each of which has a different dehydrogenation rate [1] and as such, the reaction network for \( n \)-paraffin dehydrogenation is very complicated. In addition, other reactions occur as well [2,3], examples are (i) cracking to lighter fractions, (ii) consecutive dehydrogenation and formation of aromatics via dehydrocyclization, and (iii) coke formation. These reactions take place on the metal and/or acid sites of the typically used bi-functional Pt-Sn/\( \gamma \)-Al\(_2\)O\(_3\) catalysts. In order to obtain high mono-olefin selectivity, these side reactions forming dienes, trienes/aromatics [4,5], etc., must be suppressed kinetically or inhibited by proper modification of the Pt-Sn/\( \gamma \)-Al\(_2\)O\(_3\) catalyst. The high temperatures used for \( n \)-paraffin dehydrogenation to overcome thermodynamic equilibrium limitations also results in the formation of substantial amounts of coke. Modifications of the Pt-Sn/\( \gamma \)-Al\(_2\)O\(_3\) catalyst by e.g., the introduction of (i) alkaline [6–8], alkaline earth [9–11] and transition metals [12–14], (ii) rare earth elements [15], and (iii) the use of carbon as support [16,17], have been reported to improve the mono-olefin selectivity and catalyst life-time. Recently, a Pt-Sn-K-Mg/\( \gamma \)-Al\(_2\)O\(_3\) catalyst has been commercialized in PetroChina, PR China [18], which is characterized by longer life-time (72 vs. 58 days), higher operation temperature (490 vs. 481 °C) and higher daily production (333.6 vs. 321.5 tons day\(^{-1} \)), as compared to Pt-Sn-K/\( \gamma \)-Al\(_2\)O\(_3\) catalyst. The better performance of Pt-Sn-K-Mg/\( \gamma \)-Al\(_2\)O\(_3\) catalyst was attributed to the higher mechanical strength and better thermal stability of Mg-Al-O support, as well as the moderated
acidity, enhanced interaction of Pt and support, and the increased pore volume and pore size diameter resulted from adding Mg [9]. However, detailed kinetic studies on long chain paraffin dehydrogenation are scarce, although is of high importance for determining (i) the reaction mechanism, (ii) to establish reaction networks and (iii) to be used for reactor engineering studies, e.g., for the proper simulation and design of commercial reactors.

Krylova et al. [19] investigated the kinetics of n-decane dehydrogenation over Pt-W-Li/Al₂O₃ catalyst using hydrogen/deuterium isotope exchange experiments and proposed a stepwise reaction scheme, in which the desorption of mono- and di-olefins was regarded as the rate determining step. This kinetic model was further expanded by Sadykhova et al. [20,21], and allowed determination of the kinetic parameters for n-decane, n-undecane and n-dodecane dehydrogenation over a Pt-Sn/Al₂O₃ catalyst. Basrur et al. [22] investigated the kinetics of n-decane dehydrogenation over the promoted Pt/Al₂O₃ catalyst using a Box-Williamson method and developed an empirical model for predicting paraffin conversion/olefin selectivity vs. operational parameters. Padmavathi et al. [23] studied the kinetics of n-dodecane addition, the most suitable kinetic model based on Langmuir-Hinshelwood mechanism was determined in which the surface reaction was identified as the rate determining step. This kinetic scheme was further adopted by Vafajoo et al. [24] to optimize the rate parameters for commercial plant data by a Nelder-Mead (NM) simplex method. Ivashkina et al. [25] analyzed the thermodynamics of C₆−C₁₄ dehydrogenation by means of quantum chemistry and established a kinetic model including deactivation due to coke formation. Kinetic models for deactivation of long chain paraffin dehydrogenation catalyst were also discussed by Gaidai et al. [26] and Saeedizad et al. [27]. It was shown [26,27] that coking was mainly caused by the formation of dienes. In previous studies by our group [28], it was found that three different types of coke were present at a deactivated Pt-Sn-K/Al₂O₃ catalyst used for long chain paraffin dehydrogenation. The coke was present at different positions on the catalytically active surface, viz., on Pt nanoparticle sites, acid sites on the Al₂O₃ support in close proximity of the Pt nanoparticles and discrete acid sites on the Al₂O₃ support.

It is generally accepted [29] that the dehydrogenation reactions (e.g., dehydrogenation of paraffins and mono-olefins) are catalyzed by the metal (e.g., Pt) sites of the bi-functional Pt-Sn/γ-Al₂O₃ catalyst whereas most of the side reactions (e.g., isomerization and cracking) take place on the Lewis acid sites of the catalyst. To the best of our knowledge, kinetic studies on long chain n-paraffin dehydrogenation considering both sites of the bi-functional Pt-Sn/γ-Al₂O₃ catalyst have not yet been considered, although it is highly relevant for designing an efficient catalyst with respect to mono-olefin selectivity and coking/catalyst deactivation. In this context a “bi-functional mechanistic model” is proposed for the first time. The kinetics of long chain n-paraffin dehydrogenation over an industrial Pt-Sn-K-Mg/γ-Al₂O₃ catalyst [18] using n-dodecane as the model component will be reported, applying LHHW mechanisms which were reported earlier [23,24] and commonly used for describing the catalytic dehydrogenation reactions [30]. The models include both surface reaction steps on metal nanoparticles (Pt) and acid sites. The best model was selected using appropriate model discrimination methods. Finally, relations to predict the mono- and di-olefin yield at different reaction conditions (temperature, pressure and H₂/paraffin ratio) were established and verified.

2. Experimental section

2.1. Materials

A recently commercialized Pt-Sn-K-Mg/γ-Al₂O₃ catalyst for long chain n-paraffin dehydrogenation [18], which contains 0.5 wt.% of Pt, 1.5 wt.% of Sn, 0.5 wt.% of K and 1.0 wt.% of Mg, produced at PetroChina Fushun Petrochemical Company, PR China, was used for this study. The γ-Al₂O₃ support was purchased at the Research Institute of Daily Chemical Industry, PR China. The impregnation precursors, Pt, SnCl₂, HCl, KCl, MgCl₂ and ethanol, were of analytical grade (>99.9% pure). The catalyst was prepared by wet impregnation (35 kg batch⁻¹) at 10⁻³ bar vacuum followed by drying at 70 °C (30 min), 120 °C (3 h) and calcination at 520 °C (8 h, 70 kg batch⁻¹). The catalyst was finally reduced under pure H₂ at 490 °C (8 h, 140 kg batch⁻¹). n-Dodecane (n-C₁₂, 98.88 wt.%) was commercially supplied by Liaoyang HuiFu Chemical Factory, Liaoning, PR China. It contained minor amounts of n-C₁₀ (0.05 wt.%), n-C₁₁(0.80 wt.%), n-C₁₃ (0.20 wt.% and n-C₁₄ (0.07 wt.%).

2.2. Catalyst characterization

X-ray diffraction (XRD) spectra of the catalyst were obtained using a PAN Ayalytic X’ Pert PRO instrument with Cu Kα radiation (40 kV and 40 mA) in the scan 2θ range of 20-80°.

The specific surface area of the catalyst was calculated from the adsorption isotherms obtained from nitrogen physisorption experiments at 77 K using a Micromeritics ASAP system (2010, USA) based on Brunauer-Emmett-Teller (BET) theory [31]. Pore size distributions (PSD) were calculated from the desorption branches of the isotherms according to the BJH (Barrett-Joyner-Halenda) method [32]. The catalyst was degassed at 300 °C prior to these measurements.

The total pore volume of the catalyst was measured using mercury intrusion porosimetry (MIP, Micromeritics Autopore 9520, USA). The catalyst was pre-degassed in vacuum (0.01 torr) for 1 h at 95 °C.

Scanning electron microscopy (SEM) measurements were performed

---

**Nomenclature**

<table>
<thead>
<tr>
<th>Symbol</th>
<th>Description</th>
<th>Units</th>
</tr>
</thead>
<tbody>
<tr>
<td>T</td>
<td>reaction temperature, °C</td>
<td></td>
</tr>
<tr>
<td>P</td>
<td>reaction pressure, Pa</td>
<td></td>
</tr>
<tr>
<td>Xₚₛ</td>
<td>conversion of paraffins, %</td>
<td></td>
</tr>
<tr>
<td>Yₙₒ, Yₙₜ</td>
<td>yield of mono-olefins and di-olefins, %</td>
<td></td>
</tr>
<tr>
<td>L</td>
<td>acid sites</td>
<td></td>
</tr>
<tr>
<td>M</td>
<td>Pt sites</td>
<td></td>
</tr>
<tr>
<td>Kᵢ, Kᵣ₀, Kᵣ₅, Kᵣ₆</td>
<td>equilibrium constants for reaction step i</td>
<td></td>
</tr>
<tr>
<td>Kᵣᵢ₀, Kᵣᵢ₅, Kᵣᵢ₆</td>
<td>equilibrium constants for reaction step i</td>
<td></td>
</tr>
<tr>
<td>kᵢ₋ₖᵣᵢ₁</td>
<td>reaction rate constant for reaction step i, mol h⁻¹ g⁻¹ Pa⁻¹</td>
<td></td>
</tr>
</tbody>
</table>

rₚ, rₒ, rₜ | rate of paraffins conversion, mono- and di-olefins formation, mol s⁻¹ kg⁻¹ |

Pₚ, Pₒ, Pᵢ₀, Pᵢ₅, Pᵢ₆ | partial pressure of paraffins, mono-olefins, di-olefins, aromatics and H₂, Pa |

Cᵣ₂, Cᵣ₅ | acid sites and Pt sites concentration, % |

Cᵣˢₑ, Cᵣ₆₅ | concentration of paraffins, mono-olefins, di-olefins, aromatics and H adsorbed on Pt sites, and di-olefins, aromatics adsorbed on acid sites, % |

Cᵣₒᵢ₃ₑ, Cᵣᵢ₆₅ | concentration of multifunctional intermediates C₁₃₋₁₅ and C₁₅₋₁₇ adsorbed on Pt sites, % |

W | catalyst weight, g |

Fₚ | mole flow rate of paraffins, mol h⁻¹ |

τ | W/Fₚ = space time, kg s mol⁻¹ |

m | hydrogen/paraffin mole ratio |
on a Quanta 200 (FEI Company) with an accelerating voltage of 20 kV.

Pt dispersion of the catalyst was determined by CO pulse chemisorption using a Micromeritics AutoChem II 2920 (USA). The catalyst was pre-reduced under pure H₂ (99.99%, 20 ml min⁻¹) at 500 °C for 1 h, purged with helium (99.99%, 20 ml min⁻¹) at 520 °C for 1 h and then cooled to 50 °C in He. 100 µl pulses of 5 vol% CO/He were used and the time between pulses was 4 min. The adsorbed CO was determined by TCD and Pt dispersion calculations were based on the assumption that the value of CO/Pt, is 1 [33].

2.3. Experimental setup and kinetic measurements

The experimental setup for n-dodecane dehydrogenation consisted of a micro-catalytic setup [34] with a fixed-bed reactor (tubular stainless steel, 10-mm-inner-diameter). The reactor was loaded with the commercial Pt-Sn-K-Mg/γ-Al₂O₃ catalyst (0.24 g), which was grounded to particle sizes between 0.25–0.6 mm to eliminate axial back-mixing and channeling effects in the catalyst bed. The catalyst was pre-reduced in-situ by hydrogen (99.995%, 500 ml min⁻¹) at 470 °C for 2 h. Reaction temperatures (T, 450, 460 and 470 °C) were measured by a thermocouple in the catalyst bed and used to control the reactor temperature by adjusting the electronic furnace temperature. Reactor pressures (P, 0.17, 0.24 and 0.30 MPa) were measured by a pressure gauge at the bottom of the catalyst bed and controlled by a back pressure regulator. n-Dodecane was fed to the reactor using an HPLC pump, and space times (τ = W/Fₚ) between 0.22–1.57 hcat (mol h⁻¹) were applied. The hydrogen flow was measured and controlled by mass flow controller and two hydrogen to para

The calculated values of NWP are shown in Table S1 for each experimental run. None of the experimental values exceeds 0.2, indicating the absence of internal diffusion limitation of n-dodecane. In addition, earlier experimental studies by us in the same reactor at similar reaction conditions as the present kinetic study with different catalyst particle sizes [34] also showed that internal diffusion effects were absent when using a catalyst with particle sizes in the range of 0.85–1.0 mm.

Possible internal heat transfer limitations were analyzed using the coupled concentration and temperature profiles inside the catalyst particles [35,36]. From the analysis, the calculated temperature differences inside the catalyst particles, ∆T_particle = T_surface - T_center, were found to be 0.03 K on average with a maximum value of 0.085 K (Table S1). These differences were considered too small to take into consideration.

Possible temperature differences due to external heat transfer limitations are quantified from a balance equating the rate of heat removal via the film with the rate of heat production in the catalyst particle. The results in the form of the calculated temperature differences over the film surrounding the catalyst particles, ∆T_film = T_f - T_surface, are given in Table S1 and are on average 1.5 K with a maximum value of 3.85 K. These differences were considered too small to take into consideration as well.

3. Kinetic modeling and parameter fitting for n-dodecane dehydrogenation

3.1. Reaction schemes and assumptions

Two major reaction pathways have been established for long chain paraffin dehydrogenation [2,3,37]. On an un-modified Pt/Al₂O₃ catalyst, the reaction network involves consecutive dehydrogenations to form n-olefins, n-diene and n-trienes followed by dehydrocyclization to form aromatics (RP1) [2,3,29].

Paraffins ↔ mono-olefins ↔ di-olefins ↔ tri-olefins ↔ aromatics (RP1)

In parallel, isomerization, dehydrocyclization (of paraffins), cracking and coking also take place. The latter pathway is particularly catalyzed by the acid sites of the catalyst. When the acidity of the catalyst is reduced (e.g., by alkali doping), the latter pathway is suppressed.

When using modified catalysts (Pt-Sn-K-Mg/γ-Al₂O₃ catalyst, [5]), as in this study, tri-olefins at any given time were too low for detection, in line with the observation from the industrial long chain paraffin process and the analysis of reaction products using temperature-programmed reaction/single-photon ionization time-of-flight mass spectrometry (TPRN/SP-ToF-MS) measurements [4,5], which is likely due to that they are the transient species and get converted at a very fast rate to aromatics. As such, tri-olefins are not considered in the model. Accordingly, the consecutive reaction pathway (RP1) can be simplified to RP2.

Paraffins ↔ mono-olefins ↔ di-olefins ↔ aromatics (RP2)

Langmuir - Hinselwood - Hougen - Watson (LHHW) and Power - Law (P-L) models are commonly used in heterogeneous catalysis. Padmavathi et al. [23] and Vafjoo et al. [24] have shown that LHHW
Based models fitted the experimental data well, though without specifying catalytic sites. We here consider two types of adsorption sites on the bi-functional Pt-Sn/γ-Al2O3 based catalyst, viz., metal nanoparticle sites (Pt, denoted as M) and Lewis acid sites (denoted as L). The proposed reaction scheme consists of 13 reactions, including adsorption/ desorption equilibria and surface reactions (Table 1). All reactions were considered to be reversible, elementary reactions, the only exceptions being dehydrogenation of adsorbed di-olefins on M sites (Reaction No. 8 in Table 1) and for the formation of aromatics from adsorbed di-olefins on L sites cooperated with M sites (Reaction No. 11 in Table 1). These two reactions were considered as irreversible [2,3] as thermodynamically most stable product, viz., aromatics, is formed and the backward reactions are expected to occur with very low rates. The reaction scheme is summarized in Scheme 1 and was set up using the following assumptions (H1-H10).

H1. Existence of three different active sites [28], viz., Pt nanoparticles, Al2O3 support acid sites with Pt nanoparticles in close vicinity and isolated acid sites on the Al2O3 support. As experimental techniques allow only to measure the total number of Lewis acid sites and it is not possible to estimate the fraction of the above two types of acid sites, they are lumped as Lewis acid sites to simplify the kinetic model. The Lewis acid sites of interest are a fixed fraction because there is only one catalyst being used, it will be incorporated in the rate constants for the steps where the Lewis acid sites are involved.

H2. The paraffins, mono-olefins and di-olefins react on Pt sites according to reaction sequence RP2 [5].

H3. Adsorption of paraffins occurs only on the Pt sites.

H4. Dehydrogenation of paraffins on Pt sites proceeds step-wise, viz., first formation of adsorbed M-alkyl and M−H species. In the next step, the M-alkyl species react to yield an adsorbed mono-olefin and a second M−H species.

H5. Hydrogen gas is formed by a surface reaction between two adsorbed M−H species on Pt sites [22,30].

H6. The adsorbed mono-olefins, di-olefins, aromatics and H2 species on Pt sites can be desorbed to form the corresponding products.

H7. The desorbed di-olefins can also be adsorbed on the Lewis acid sites. The desorbed mono-olefins from metal sites was not considered to re-adsorb on Lewis acid sites due to the adsorption on Lewis acid sites is always weaker than metal sites and minor. And practically, the catalyst design and reaction conditions (e.g., low contact time) for long chain paraffin dehydrogenation are optimized to render the mono-olefins desorption a facile process.

H8. The adsorbed di-olefins on Lewis acid sites can be transformed into aromatics via dehydrocyclization in cooperation of Pt sites [29] follows step-wise reaction mechanisms. These stepwise reactions are lumped in one reaction to simplify the model, due to the fact that dehydrocyclization is not the core reaction and minimized in long chain paraffin dehydrogenation process.

H9. The adsorbed aromatics on the Lewis acid sites can be desorbed to form aromatics.

H10. Coking and cracking reactions are not included in the model. This is due to the fact that their reaction rates are so low that the corresponding products are not detectable at the reaction conditions applied. This can also be indicated by the reported long life-time of and low coke deposition on the catalysts for long chain n-paraffins dehydrogenation, which were 30 days and 8 wt.% for Pt-Sn-Li/Al2O3 [38], 37 days and 7.9 wt.% for Pt-Sn-In-Fe/Al2O3 [39], and 72 days and 3.8 wt.% for the Pt-Sn-K-Mg/γ-Al2O3 catalyst investigated in this kinetic study [18].

3.2. From reaction network to kinetic models

To obtain expressions for the three reaction rates rP, rO and rD (RP2), the Langmuir-Hinshelwood-Hougen-Watson (LHHW) approach was followed, assuming that certain reactions among those presented in Table 1 are rate determining (rds). A total of 6 different kinetic models were derived:

3.2.1. Model 1: Adsorption of paraffins (reaction 1 in Table 1) is rate determining

\[
\begin{align*}
    r_P &= k_P P_D - \frac{k_{DM} P_D}{k_{DP}} (K_D P_D)^3 \\
    r_O &= k_0 O_D - \frac{k_{OM} O_D}{k_{MO}} k_{DP} P_D \\
    r_D &= k_D' D_M \frac{K_{DM} P_D}{D_M} + k_D'' D_M \frac{K_{DP} P_D}{(1 + K_{DP} P_D + K_{DL} P_D) D_M}^4
\end{align*}
\]

Table 1

<table>
<thead>
<tr>
<th>Reaction steps</th>
<th>Elementary reaction</th>
<th>Reaction rate</th>
<th>Equilibrium constant</th>
</tr>
</thead>
<tbody>
<tr>
<td>1. Paraffins adsorption on M sites</td>
<td>C12H28 + M ↔ C12H24M</td>
<td>( k_1 (P_M P_D - \frac{K_{DP}}{K_{DM}} P_D) )</td>
<td>( K_P = \frac{C_{PM}}{P_M C_{PM}} )</td>
</tr>
<tr>
<td>2. Dehydrogenation of adsorbed paraffins on M sites – 1st step</td>
<td>C12H24M + M ↔ C12H20M + HM</td>
<td>( k_2 (C_{OM} C_M - \frac{1}{K_{OM}} C_{OM} C_{TM}) )</td>
<td>( K_2 = \frac{C_{PM} C_{TM}}{C_{OM} C_{TM}} )</td>
</tr>
<tr>
<td>3. Dehydrogenation of adsorbed paraffins on M sites – 2nd step</td>
<td>C12H20M + M ↔ C12H16M + HM</td>
<td>( k_3 (C_{OM} C_M - \frac{1}{K_{OM}} C_{OM} C_{TM}) )</td>
<td>( K_3 = \frac{C_{PM} C_{TM}}{C_{OM} C_{TM}} )</td>
</tr>
<tr>
<td>4. Mono-olefins desorption</td>
<td>C12H16M + M ↔ C12H12M + M</td>
<td>( k_4 (C_{OM} C_M - \frac{1}{K_{OM}} C_{OM} C_{TM}) )</td>
<td>( K_4 = \frac{C_{PM} C_{TM}}{C_{OM} C_{TM}} )</td>
</tr>
<tr>
<td>5. Dehydrogenation of adsorbed mono-olefins on M sites – 1st step</td>
<td>C12H12M + M ↔ C12H8M + HM</td>
<td>( k_5 (C_{OM} C_M - \frac{1}{K_{OM}} C_{OM} C_{TM}) )</td>
<td>( K_5 = \frac{C_{PM} C_{TM}}{C_{OM} C_{TM}} )</td>
</tr>
<tr>
<td>6. Dehydrogenation of adsorbed mono-olefins on M sites – 2nd step</td>
<td>C12H8M + M ↔ C12H4M + HM</td>
<td>( k_6 (C_{OM} C_M - \frac{1}{K_{OM}} C_{OM} C_{TM}) )</td>
<td>( K_6 = \frac{C_{PM} C_{TM}}{C_{OM} C_{TM}} )</td>
</tr>
<tr>
<td>7. Di-olefins desorption</td>
<td>C12H6M + M ↔ C12H4M + M</td>
<td>( k_7 (C_{OM} C_M - \frac{1}{K_{OM}} C_{OM} C_{TM}) )</td>
<td>( K_7 = \frac{C_{PM} C_{TM}}{C_{OM} C_{TM}} )</td>
</tr>
<tr>
<td>8. Dehydrogenation of adsorbed di-olefins on M sites</td>
<td>C12H4M + 4M ↔ C12H2M + 4HM</td>
<td>( k_8 (C_{OM} C_M - \frac{1}{K_{OM}} C_{OM} C_{TM}) )</td>
<td>( K_8 = \frac{C_{PM} C_{TM}}{C_{OM} C_{TM}} )</td>
</tr>
<tr>
<td>9. Di-olefins adsorption on L site</td>
<td>C12H4M + L ↔ C12H2L</td>
<td>( k_9 (P_{LM} P_D - \frac{K_{DL}}{K_{DL}} P_D) )</td>
<td>( K_{DL} = \frac{C_{DL}}{P_{LM} C_{DL}} )</td>
</tr>
<tr>
<td>10. Aromatics desorption from M site</td>
<td>C12H8M + M ↔ C12H4M + M</td>
<td>( k_{10} (C_{OM} C_M - \frac{1}{K_{OM}} C_{OM} C_{TM}) )</td>
<td>( K_{OM} = \frac{C_{OM} C_{TM}}{C_{OM} C_{TM}} )</td>
</tr>
<tr>
<td>11. Aromatics formation from adsorbed di-olefins on L sites cooperated with M sites</td>
<td>C12H6L + 4M ↔ C12H2L + 4HM</td>
<td>( k_{11} (C_{OM} C_M - \frac{1}{K_{OM}} C_{OM} C_{TM}) )</td>
<td>( K_{OM} = \frac{C_{OM} C_{TM}}{C_{OM} C_{TM}} )</td>
</tr>
<tr>
<td>12. Aromatics desorption from L sites</td>
<td>C12H4L + L ↔ C12H2L</td>
<td>( k_{12} (C_{OM} C_M - \frac{1}{K_{OM}} C_{OM} C_{TM}) )</td>
<td>( K_{OM} = \frac{C_{OM} C_{TM}}{C_{OM} C_{TM}} )</td>
</tr>
</tbody>
</table>

L: Lewis acid sites; M: Pt sites.
3.2.2. Model 2: Dehydrogenations of absorbed paraffins and mono-olefins (surface reactions 2 and 5 in Table 1) are the rds

\[
\begin{align*}
  r_p &= k_p' \frac{K_P P_R (K_H P_R)^{0.5}}{K_D P_D} \\
  r_o &= k_o' \frac{K_O P_O - \frac{K_O P_O}{K_K} (K_H P_R)^{0.5}}{K_D P_D} \\
  r_D &= k_D' \frac{K_D P_D}{DEHN^2} + k_{DL} \frac{K_D P_D}{(1 + K_D P_D + K_A P_A)DEHN^2}
\end{align*}
\]

\[
DHN = 1 + K_P P_R + \frac{K_O P_O}{K_K} (K_H P_R)^{0.5} + K_O P_O + \frac{K_D P_D}{K_D} (K_H P_R)^{0.5} + K_D P_D + K_A M P_A + (K_H P_R)^{0.5}
\]

3.2.3. Model 3: Dehydrogenations of absorbed paraffins and mono-olefins (surface reactions 3 and 5 in Table 1) are the rds

\[
\begin{align*}
  r_p &= k_p' \frac{K_P K_P P_R (K_H P_R)^{0.5}}{K_D P_D} \\
  r_o &= k_o' \frac{K_O P_O - \frac{K_O P_O}{K_K} (K_H P_R)^{0.5}}{K_D P_D} \\
  r_D &= k_D' \frac{K_D P_D}{DEHN^3} + k_{DL} \frac{K_D P_D}{(1 + K_D P_D + K_A P_A)DEHN^3}
\end{align*}
\]

\[
DHN = 1 + K_P P_R + K_K K_P P_R (K_H P_R)^{0.5} + K_O P_O + \frac{K_D P_D}{K_D} (K_H P_R)^{0.5} + K_D P_D + K_A M P_A + (K_H P_R)^{0.5}
\]

3.2.4. Model 4: Dehydrogenations of absorbed paraffins and mono-olefins (surface reactions 2 and 6 in Table 1) are the rds

\[
\begin{align*}
  r_p &= k_p' \frac{K_P P_R - \frac{K_O P_O}{K_K} (K_H P_R)^{0.5}}{DEHN^2} \\
  r_o &= k_o' \frac{K_O P_O - \frac{K_O P_O}{K_K} (K_H P_R)^{0.5}}{DEHN^2} \\
  r_D &= k_D' \frac{K_D P_D}{DEHN^2} + k_{DL} \frac{K_D P_D}{(1 + K_D P_D + K_A P_A)DEHN^2}
\end{align*}
\]

\[
DHN = 1 + K_P P_R + \frac{K_O P_O}{K_K} (K_H P_R)^{0.5} + K_O P_O + K_K K_P P_R (K_H P_R)^{0.5} + K_O P_O + K_A M P_A + (K_H P_R)^{0.5}
\]

3.2.5. Model 5: Dehydrogenations of absorbed paraffins and mono-olefins (surface reactions 3 and 6 in Table 1) are the rds

\[
\begin{align*}
  r_p &= k_p' \frac{K_P K_P P_R (K_H P_R)^{0.5}}{K_D P_D} \\
  r_o &= k_o' \frac{K_O P_O - \frac{K_O P_O}{K_K} (K_H P_R)^{0.5}}{K_D P_D} \\
  r_D &= k_D' \frac{K_D P_D}{DEHN^2} + k_{DL} \frac{K_D P_D}{(1 + K_D P_D + K_A P_A)DEHN^2}
\end{align*}
\]

\[
DHN = 1 + K_P P_R + K_K K_P P_R (K_H P_R)^{0.5} + K_O P_O + K_K K_P P_R (K_H P_R)^{0.5} + K_O P_O + K_A M P_A + (K_H P_R)^{0.5}
\]

3.2.6. Model 6: Desorption of mono-olefins and di-olefins (reactions 4 and 7 in Table 1) are the rds

\[
\begin{align*}
  r_p &= k_p' \frac{K_P K_P K_P P_R (K_H P_R)^{0.5}}{K_D P_D} \\
  r_o &= k_o' \frac{K_O P_O - \frac{K_O P_O}{K_K} (K_H P_R)^{0.5}}{K_D P_D} \\
  r_D &= k_D' \frac{K_D P_D}{DEHN^2} + k_{DL} \frac{K_D P_D}{(1 + K_D P_D + K_A P_A)DEHN^2}
\end{align*}
\]

\[
DHN = 1 + K_P P_R + K_K K_P P_R (K_H P_R)^{0.5} + K_O P_O + K_K K_P P_R (K_H P_R)^{0.5} + K_O P_O + K_A M P_A + (K_H P_R)^{0.5}
\]
The dependency of kinetic parameters and equilibrium constants on temperature can be indicated by Arrhenius equation (Eq. 5) and Van’t Hoff equation (Eq. 6), separately.

$$k_i = k_{0,i} \exp\left(-\frac{E_{a,i}}{RT}\right)$$  \hspace{1cm} (5)

$$K_i = K_{0,i} \exp\left(-\frac{\Delta H_{i}}{RT}\right)$$  \hspace{1cm} (6)

Hence, for each parameter stated in the formulae above, two kinetic constants were estimated, viz., the pre-exponential factor ($k_{0,i}$ or $K_{0,i}$) and the activation energy ($E_{a,i}$) or enthalpy ($\Delta H_{i}$). Most parameters were imposed to be positive, in order to be physically meaningful. However, the enthalpy of adsorption reactions (e.g., for reactions No. 1, 4, 7, 10 and 12, Table 1) were set to be negative because adsorption is an exothermic process.

3.3. Reactor modeling and parameter estimation

The reactions were carried out in a fixed bed reactor and it was assumed that the reactor behaves as a PFR reactor. Volume expansion in the reactor was not considered as it is limited due to the low conversion of n-dodecane (e.g., < 10%) and very high amount of hydrogen gas (e.g., $H_2$/paraffin mole ratios was 3:1 - 6:1) was used as the dilution gas. As such, the following 5 ordinary differential equations (ODEs, Eqs. 7-11) were used for reactor modeling.

$$\frac{dP_T}{d\tau} = -P_T \rho_f$$  \hspace{1cm} (7)

$$\frac{dP_i}{d\tau} = P_T (r_i - r_0)$$  \hspace{1cm} (8)

$$\frac{dP_D}{d\tau} = P_T (r_0 - r_D)$$  \hspace{1cm} (9)

$$\frac{dP_{H_2}}{d\tau} = P_T (r_D + 2r_0)$$  \hspace{1cm} (10)

$$\frac{dP_{H_2}}{d\tau} = P_T (r_D + 2r_0)$$  \hspace{1cm} (11)

where $P_T$ is the total pressure and $\tau$ is the space time. The initial conditions are represented by the initial partial pressures of n-dodecane and hydrogen, which can be calculated from the experimental pressure and the hydrogen to paraffin ratio in the feed (m). By integrating the system between 0 and the desired space time $\tau$, it is possible to determine the partial pressures of all the species involved. Eqs. 7-11 were implemented in the software package Matlab™ (The Mathworks, Inc.) and solved using the function ode15s.

Parameter estimation was performed in Matlab™ using the function lsqnonlin, which is based on a non-linear least square minimization method and involved all 60 experimentally obtained data points (Table S1). The sum of the normalized squared deviations (NSD) was optimized. For the $i$th species and the $j$th experiment, the normalized deviation $NSD_{ij}$ is defined as Eq. 12.

$$NSD_{ij} = \left(\frac{X_{ij} - X_{ij}^{\text{exp}}}{X_{ij}^{\text{exp}}}ight)^2$$  \hspace{1cm} (12)

where $x$ is the conversion (for paraffins) or yield (for mono-olefins and di-olefins) for model and experiments. The use of normalized deviations was preferred over relative errors as the latter method gives an excessive weight to the smaller values, resulting in a poor fitting for the higher ones.

For model discrimination, the root mean squared error (RMSE) and the Pearson’s correlation coefficient $R^2$ for each model were calculated for each species.

4. Results and discussions

4.1. Characteristics of the industrial Pt-Sn-K-Mg/γ-Al2O3 catalyst

The Pt-Sn-K-Mg/γ-Al2O3 catalyst was characterised using XRD, nitrogen physisorption, mercury intrusion porosimetry (MIP) and SEM and the results are shown in Fig. 1 and summarised in Table 2. XRD patterns of the catalyst (Fig. 1a) show h-k-l reflections characteristic of the γ-Al2O3 phase (JCPD No. 04-0858). The nitrogen adsorption/desorption data (Fig. 1b) of the catalyst show a type IV isotherm and a H1-type hysteresis loop [40], indicating the presence of mesoporous and macroporous cylindrical pores. This is further illustrated by the PSD obtained by the BJH method (Fig. 1b) and the MIP data (Fig. 1c). The bimodal PSD structure (Fig. 1c, centered at 14.6 nm and 1300 nm) is typical for long chain paraffin dehydrogenation catalysts.

Fig. 1. (a) XRD, (b) BET, (c) MIP and (d) SEM of Pt-Sn-K-Mg/γ-Al2O3 catalyst.
4.2. Effect of the process conditions on n-dodecane dehydrogenation

The yields of mono-olefins and di-olefins during n-dodecane dehydrogenation over Pt-Sn-K-Mg/γ-Al₂O₃ catalyst at different temperatures (T), pressures (P), H₂/paraffin ratios (m) and space times (τ) are plotted in Fig. 2. It can be seen from Fig. 2 that when P, m and τ are constant (e.g., P = 0.24 MPa, m = 6:1 and τ = 0.45), higher reaction temperature leads to an increase in the yields of mono-olefin (Y₀, Fig. 2a, c and e) and di-olefin (Y₀, Fig. 2b, d and f). When T, m and τ are the same (e.g., T = 450 °C, m = 6:1 and τ = 0.45), increasing the reaction pressure decreases the yields of mono-olefin (Y₀, Fig. 2a) and di-olefin (Y₀, Fig. 2b). When T, P and τ are same (e.g., T = 460 °C, P = 0.17 MPa and τ = 0.45), increasing the H₂/paraffin ratio (m) lowers the yields of mono-olefin (Y₀, Fig. 2c) and di-olefin (Y₀, Fig. 2d). When T, P and m are constant (e.g., T = 470 °C, P = 0.30 MPa and m = 6:1), increasing the space time (τ) over the catalyst favors higher yields of mono-olefin (Y₀, Fig. 2e) and di-olefin (Y₀, Fig. 2f). These trends follow from the thermodynamics of dehydrogenation of long chain paraffins to olefins, which is an endothermic reversible reaction accompanied by volume expansion.

The effect of the space time (τ) and yields of mono- and di-olefins (Y₀ and Y₀) was modeled using a simple empirical approach (Eq. 13).

\[ Y = -A \times \exp \left( \frac{-C}{B} \right) \]  

(13)

The results are shown in Fig. 2 (solid lines) and the equations with parameter values are given in the Supplementary Information. Agreement between the experimental data and the predicted values is very good. The Empirical modeling (Eqs. 13, S1 and S2) and our kinetic testing (Fig. 2) reveal that the increased selectivity of mono-olefins (or the ratio of mono-olefins yields to di-olefins yields) can be obtained by operating at low space time (when P, T and m are same), low reaction temperature (when τ, P and m are same), high pressure (when τ, T and m are same) and high H₂/paraffin ratio (when τ, P and T are same).

### Table 2

Characteristics of the Pt-Sn-K-Mg/γ-Al₂O₃ catalyst used in this study.

<table>
<thead>
<tr>
<th>Catalysts</th>
<th>Shape</th>
<th>Bulk density (g cm⁻³)</th>
<th>S_{BET}(N₂) (m² g⁻¹)</th>
<th>Pore volume (cm³ g⁻¹)</th>
<th>Pt dispersion (%)</th>
</tr>
</thead>
<tbody>
<tr>
<td>Pt₀.₅Sn₁.₅K₀.₅Mg₁.₀/γ-Al₂O₃</td>
<td>Granule, φ1.25 – 2.5 mm</td>
<td>0.33</td>
<td>149</td>
<td>1.46</td>
<td>70</td>
</tr>
</tbody>
</table>

* nominal metal loading, wt.%.

**Fig. 2.** Mono-olefin and di-olefin yield for n-dodecane dehydrogenation over Pt-Sn-K-Mg/γ-Al₂O₃ catalyst at different reaction conditions.
sites to the adsorbed di-olefin—these steps seem to be the rate-determining steps. The models based on adsorption (Model 1) and desorption (Model 6) as the rds show significant differences. The models based on surface reactions (Models 2–5) are consistent with previous research. This is also supported by considering the parity plots in Fig. 4 and Figs. S2-S5.

At first sight, the model quality indicators for the remaining four models with surface reactions as the rds are rather similar (Table 3). However, Model 4 and 5 present slightly lower values for the RMSE and NSD compared to the other two surface-reaction-based models (Models 2 and 3). Both Models 4 and 5 assume that the rds is the dehydrogenation of half-dehydrogenated C12H23M species adsorbed on Pt sites to the adsorbed di-olefin, viz., the second step of the mono-olefin dehydrogenation to di-olefin (Reaction 6, Table 1). These findings imply that the second step of the mono-olefin dehydrogenation is slower than the first.

The activation energies of Model 4 (54.5 KJ mol⁻¹) and Model 5 (59.3 KJ mol⁻¹) for dodecane dehydrogenation to mono-olefins are in the range (31.4–78.0 KJ mol⁻¹) reported for dodecane dehydrogenation by Sadykhova et al. [21], Kang et al. [41], Padmaavathi et al. [23] and Jiang et al. [42]. Further detailed comparison between Models 4 and 5 shows that the Pearson’s correlation coefficient for Model 5 is better than for Model 4, in line with the parity plots in Fig. 4. Thus Model 5 was selected as the best model for this kinetic study, which assumed that the rate determining steps are represented by the second step in the dehydrogenation of both paraffin and mono-olefin (rP and rD).

The activation energies of Model 5 for dodecane dehydrogenation to mono-olefins (59.3 KJ mol⁻¹) and for mono-olefins dehydrogenation to di-olefins (53.9 KJ mol⁻¹) are very close, indicating the formation of mono-olefins and di-olefins have similar sensitivity with the temperature increase. This is also well reflected by the small change of mono-olefins to di-olefins ratio when the temperature increases from 450 °C to 470 °C during the kinetic testing. Practically, we have also observed that the selectivity of mono-olefins is relatively stable (or slightly decreased) during the life-span of 72 days when the operation temperature in the industrial plant increases from 478 °C to 490 °C to keep the productivity [18]. This can be kinetically explained from our kinetic modeling which proves that the increase of temperature (e.g., 450–490 °C) has little effect on mono-olefins to di-olefins ratio related to their similar reaction activation energy. As such, other operation parameters, e.g., pressure, should be resorted to more effectively tune the mono-olefin selectivity for long chain paraffins dehydrogenation, indicated by the above kinetic testing and modeling.

A previous mechanistic study [29] has shown that the metal sites are required to cleave the C–H bond of the adsorbed paraffin (or mono-olefin). Only when the two neighboring C atoms are activated, the corresponding mono-olefin (or di-olefin) can be formed (RP3). The present kinetic results reveal that the half-dehydrogenated paraffin and mono-olefin adsorbed on the metal sites are more difficult to be dehydrogenated (Step 2, Reactions No. 3 and 6, Table 1) on the neighboring metal sites than the first step (dehydrogenation of the paraffin and mono-olefin adsorbed on the metal sites, Reactions No. 2 and 5, Table 1). This is due to the fact that the second dehydrogenation step requires extra metal sites which are present and adjacent to those adsorbing the half-dehydrogenated species. Thus lowering the liquid hourly space velocity (LHSV) of paraffins (viz., increasing T) while keeping P and m same can decrease the concentration of the adsorbed paraffin (or mono-olefins) on metal sites (e.g., Pt) surface, leading to the increased yields of mono-olefins and di-olefins which is in line with the observation from our kinetic testing (Fig. 2). Alternatively, increasing the loading of the noble metal (e.g., Pt) on the long chain paraffin dehydrogenation catalyst might be another option to increase the overall reaction rate. However, it was often observed that increasing Pt loading decreases the Pt dispersion [43], resulting in a lowering of the rate of olefin formation [44]. Our kinetic results also indicate that the noble metals (e.g., Pt) are not required in large quantities but are needed to be highly dispersed on the support (e.g., γ-Al2O3 with high surface area) to generate high concentration of Pt particle surface per support surface area.

It is interesting to notice that the reaction rate for the aromatization step (Table 1) for the best model (Model 5) includes two additive terms. One involves KOH, which takes into account the synergy between Pt and acid sites (Reaction 11, Table 1) and the other one is KOM, where the reaction conditions were used for kinetic modeling. The model quality indicators (NSD, RMSE and R²) for the 6 different models tested are provided in Table 3 whereas the corresponding kinetic parameters for each model are given in Table 4. The models based on adsorption (Model 1) and desorption (Model 6) as the rds show significantly higher values for the NSD and RMSE and lower values for R² than the models based on surface reactions (Models 2–5). As such, models with surface reactions as the rate determining steps seem to fit the experimental data better, which is consistent with previous research [23]. This is also supported by considering the parity plots in Fig. 4 and Figs. S2-S5.

**Table 3**

<table>
<thead>
<tr>
<th>Reaction model</th>
<th>Normalized squared deviation (NSD)</th>
<th>Root of mean squared error (RMSE)</th>
<th>Pearson’s correlation coefficient (R²)</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>P</td>
<td>O</td>
<td>D</td>
</tr>
<tr>
<td>Model 1</td>
<td>0.3392</td>
<td>0.0612</td>
<td>0.1388</td>
</tr>
<tr>
<td>Model 2</td>
<td>0.2257</td>
<td>0.0726</td>
<td>0.0890</td>
</tr>
<tr>
<td>Model 3</td>
<td>0.2591</td>
<td>0.0623</td>
<td>0.1149</td>
</tr>
<tr>
<td>Model 4</td>
<td>0.2218</td>
<td>0.0795</td>
<td>0.0631</td>
</tr>
<tr>
<td>Model 5</td>
<td>0.2213</td>
<td>0.0881</td>
<td>0.0615</td>
</tr>
<tr>
<td>Model 6</td>
<td>1.0703</td>
<td>0.1497</td>
<td>0.0806</td>
</tr>
</tbody>
</table>

P: paraffins; O: mono-olefins; D: di-olefins.
Table 4
Reaction rate constants, reaction equilibrium constants and adsorption constants for n-dodecane dehydrogenation over Pt-Sn-K-Mg/γ-Al₂O₃ catalysts. The values of activation energy are given in kJ mol⁻¹.

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</thead>
<tbody>
<tr>
<td>1</td>
<td>4.1E-6 × exp(1.0 RT) Pa⁻¹</td>
<td>4.5E-6 × exp(1.2 RT) Pa⁻¹</td>
<td>3.0E-6 × exp(1.9 RT) Pa⁻¹</td>
<td>4.6E-6 × exp(1.2 RT) Pa⁻¹</td>
<td>3.0E-6 × exp(1.6 RT) Pa⁻¹</td>
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pathway only involves the reaction on Pt sites (Reaction No. 8, Table 1). The estimated kinetic parameters in Table 4, along with the simulation results, indicate that the former pathway is predominant. Thus, the above calculations prove the reported assumption that aromatization and further coking are accelerated by acid sites together with Pt sites [2]. Because highly dispersed metal (Pt) sites are required for high reaction rates for long chain paraffin dehydrogenation to olefins, neutralization of the Lewis acid sites is expected to lower aromatics and coke formation rates for improved olefin selectivity and prolonged catalyst lifetime. Indeed, this is done in practice by modifying the catalysts with alkali metals such as Li [7,8], Na [45], K [6,7], and Cs [45].

5. Conclusions

The empirical model presented here for the kinetics of n-dodecane dehydrogenation over the Pt-Sn-K-Mg/γ-Al₂O₃ catalyst shows promise by successfully predicting the yields of mono- and di-olefins within the industrially relevant conditions (450–470 °C, 0.17 – 0.30 MPa, H₂/paraffin mole ratios between 3:1 and 6:1, space times between 0.22–1.57 g h mol⁻¹). Operation at low space time, high pressure, high H₂/paraffin ratio and low reaction temperature is favorable to high mono-olefins selectivity. The bi-functional model is relevant because it is based on the elementary reactions on Pt and Lewis acid sites, key characteristics of this dehydrogenation catalyst. Within the six reaction rate models tested, the surface step involving the interaction of Pt-H and Pt-alkyl species to result in an absorbed mono-olefin or di-olefin is observed to be the kinetically relevant rate determining step. The kinetic modeling indicates that the increased temperature within a certain range (e.g., 450–490 °C) has little effect on mono-olefin selectivity, which opens the opportunity for the industry to maintain the productivity at later period of catalyst life-time by increasing the operation temperature while keeping relatively stable product selectivity. Kinetic data also suggests that aromatization was majorly caused by Pt and Lewis acid sites in a concerted way. This is critical in the design of an efficient catalyst as, lowering the aromatics formation by neutralizing Lewis acid sites can (i) improve olefin selectivity and (ii) prolong catalyst life-time as aromatics are precursor to coke. Both are extremely relevant of commercial operation of long chain paraffins dehydrogenation catalysts.

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Fig. 4. Fitting results of Model 4 (a) and Model 5 (b) versus experimental results.
&D and commercialization activities regarding paraffin dehydrogenation and Y. Lai for his assistance on kinetic measurements.

Appendix A. Supplementary data

Supplementary material related to this article can be found, in the online version, at doi:https://doi.org/10.1016/j.apcata.2019.04.026.

References