Catalytic Hydrotreatment of Triglycerides and Free Fatty Acids for Green Diesel Production

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Unit of Hydrocarbons and Biofuels Processing
• Introduction
• Laboratory Research
  ➢ HDO reactions
  ➢ Equipment
  ➢ Conventional HDS in the presence of V.O.
  ➢ Deep HDS in the presence of V.O.
• Industrial Reactor Simulation
• Conclusions
V.O./Fats Conversion to Biodiesel (FAME)

Main Process / Product Characteristics

- New production equipment
- Consumption of catalyst and methanol
- Product compliance with standards
- By-product glycerin
- Biodiesel oxidation during storage

Direct Use of V.O. in diesel engines

- Not recommended
- Use of refined VO
- Long-term use causes problems in engine parts (gumming-up of injectors and coking-up of valves and exhaust)
## V.O./Fats Conversion to Green Diesel

**Main Process / Product Characteristics**

- Use of existing equipment in refineries
- Known technology – Hydrotreatment
- Use of conventional hydrotreating catalysts
- Product fully compatible with diesel
- Product cetane number up to 85
- Storage stability

## Processing of V.O./Fats in Refineries

- **GOAL**
  
  Replacement of part of petro-diesel with renewable diesel produced in a refinery via an existing process

- **Target**
  
  The prediction of the operation of Industrial Units fed with Gas Oil /V.O. mixtures
  
  The suggestion of the necessary modifications to the existing Units for successful operation
**Liquid Phase**

- **Decarboxylation**
  \[
  \text{Triglyceride} + 4 \text{H}_2 \rightarrow \text{Green Diesel} + \text{CO}_2 + \text{C}_3\text{H}_8 + 4 \text{H}_2
  \]

- **Decarbonylation**
  \[
  \text{Triglyceride} + 2 \text{H}_2 \rightarrow \text{Green Diesel} + \text{CO} + \text{C}_3\text{H}_8 + \text{H}_2\text{O}
  \]

- **Hydrogenation/Dehydration**
  \[
  \text{Triglyceride} + \text{H}_2 \rightarrow \text{Green Diesel} + \text{C}_3\text{H}_8 + 2 \text{H}_2\text{O}
  \]

**Gas Phase**

- \[
  \text{CO}_2 + \text{H}_2 \leftrightarrow \text{CO} + \text{H}_2\text{O}
  \]

- \[
  \text{CO}_2 + 4\text{H}_2 \rightarrow \text{CH}_4 + 2\text{H}_2\text{O}
  \]

- \[
  \text{C}_3\text{H}_8 + \text{H}_2\text{O} \rightarrow \text{CO} + 2\text{H}_2\text{O}
  \]
- F: Liquid feed tank
- P: Pump
- R: Reactor
- S: Gas - Liquid separator
- WGM: Wet Gas Meter
- Red Lines : Upflow Operation

1 March 2013
Catalyst: 5 g CoMo/γ-Al2O3 (Albemarle KF 757)

- Inert
- Glass wool
- Diluted Catalytic Bed
- Glass wool
- Inert

Laboratory Reactor

UGent/FCh13/3L

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Experimental Conditions

- Liquid Feed: 10% wt Cotton Seed Oil in Diesel (S < 20 ppm).
- Reaction Temperatures: 305 – 320 – 335 °C
- Pressure: 30 bar
- Liquid feed: 25-125 g/h, WHSV: 5-25 g_{oil}/g_{cat}/h
- Gas feed: 24 Nl_{H2}/h
- Gas composition: 1% vol/vol H2S / H2 (DMDS)

Analysis

- Saponification Number (ASTM D94-02)

<table>
<thead>
<tr>
<th>Fatty Acids Type</th>
<th>Form</th>
<th>Formula</th>
<th>% (kg/kg)</th>
</tr>
</thead>
<tbody>
<tr>
<td>Palmitic</td>
<td>16:0</td>
<td>C_{16}H_{32}O_{2}</td>
<td>22.96</td>
</tr>
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<td>C_{16}H_{30}O_{2}</td>
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<td>18:0</td>
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<td>2.30</td>
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<td>Oleic</td>
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<td>C_{18}H_{32}O_{2}</td>
<td>55.40</td>
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<tr>
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<td>18:2</td>
<td>C_{18}H_{32}O_{2}</td>
<td>0.02</td>
</tr>
<tr>
<td>Linolenic</td>
<td>18:3</td>
<td>C_{18}H_{30}O_{2}</td>
<td>0.20</td>
</tr>
</tbody>
</table>
HDO Catalytic Reaction rates

\[-r_{HDO} = k_{HDO} \cdot C_{est_0} \cdot (1 - x)\]

\[
\text{WHSV} = \frac{Q_L}{V_{\text{cat}}}
\]

T=305 °C

T=335 °C

T=320 °C

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Estimated Parameters

\[ k_{\text{HDO},0} \quad \text{kg}_{\text{feed}}/\text{hr}/\text{kg}_{\text{cat}} \]

\[ E_{\text{HDO}} \quad \text{kJ}/\text{mol} \]

3.18 \times 10^{10} \quad 108
<table>
<thead>
<tr>
<th>Properties</th>
<th>Diesel</th>
<th>Diesel/GD, 9/1</th>
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<tbody>
<tr>
<td>Density (g/ml)</td>
<td>0.8533</td>
<td>0.8497</td>
</tr>
<tr>
<td>Cetane Number</td>
<td>50.6</td>
<td>53.6</td>
</tr>
<tr>
<td>Cetane Index</td>
<td>51.8</td>
<td>53.9</td>
</tr>
</tbody>
</table>

Distillation Curve

- : Desulphurized Diesel
- : Diesel / Green Diesel 9/10
HDS / HDO Set 1

Experimental Conditions

- Liquid Feed: Gasoil (0.88% S) / Palm Oil (0, 2.5, 5.0 and 10.0 % wt) mixtures:
- Catalyst: 40 g CoMo/γ-Al$_2$O$_3$ (Albemarle KF 757), diluted with 50 g SiC.
- Reaction Temperature: 310 – 330 – 350 °C
- Reactor Pressure: 33 bar
- Liquid flow rate: 56, 40 and 28 g/h, WHSV: 1.4, 1.0 and 0.7 g$_{\text{oil}}$/g$_{\text{cat}}$/h
- Gas flow rate: 20 Nl$_{\text{H}_2}$/h

Analysis

- Sulphur content in product oil, Exit gas composition
Liquid feeds properties

Palm Oil composition

<table>
<thead>
<tr>
<th>Fatty Acids</th>
<th>Form</th>
<th>Formula</th>
<th>% wt</th>
</tr>
</thead>
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<tr>
<td>Lauric</td>
<td>12:0</td>
<td>C_{12}H_{24}O_{2}</td>
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<tr>
<td>Myristic</td>
<td>14:0</td>
<td>C_{14}H_{28}O_{2}</td>
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<tr>
<td>Palmitic</td>
<td>16:0</td>
<td>C_{16}H_{32}O_{2}</td>
<td>44.3</td>
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<td>Palmitoleic</td>
<td>16:1</td>
<td>C_{16}H_{30}O_{2}</td>
<td>0.2</td>
</tr>
<tr>
<td>Stearic</td>
<td>18:0</td>
<td>C_{18}H_{36}O_{2}</td>
<td>5.0</td>
</tr>
<tr>
<td>Oleic</td>
<td>18:1</td>
<td>C_{18}H_{34}O_{2}</td>
<td>39.1</td>
</tr>
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<td>Linoleic</td>
<td>18:2</td>
<td>C_{18}H_{32}O_{2}</td>
<td>10.1</td>
</tr>
<tr>
<td>Linolenic</td>
<td>18:3</td>
<td>C_{18}H_{30}O_{2}</td>
<td>0.1</td>
</tr>
</tbody>
</table>

Gas Oil Properties

<table>
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<tr>
<th>Property</th>
<th>Value</th>
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<tbody>
<tr>
<td>Density (g/ ml)</td>
<td>0.865</td>
</tr>
<tr>
<td>Water (ppm)</td>
<td>250</td>
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<tr>
<td>Sulphur (% w/ w)</td>
<td>0.88</td>
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<tr>
<td>Cetane Number</td>
<td>48.3</td>
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</table>

Gasoil Distillation Curve (% -°C)

<table>
<thead>
<tr>
<th>IBP</th>
<th>5%</th>
<th>10%</th>
<th>20%</th>
<th>30%</th>
<th>40%</th>
<th>50%</th>
<th>60%</th>
<th>70%</th>
<th>80%</th>
<th>85%</th>
<th>90%</th>
<th>95%</th>
<th>FBP</th>
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</thead>
<tbody>
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<td></td>
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<tr>
<td>174</td>
<td>233</td>
<td>248</td>
<td>264</td>
<td>274</td>
<td>284</td>
<td>294</td>
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<td>331</td>
<td>338</td>
<td>348</td>
<td>360</td>
<td>368</td>
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</tbody>
</table>
Experimental Results for Set 1

T=310 C

- Cs ppm
- % VEGETABLE OIL

- WHSV=0.7 h⁻¹
- WHSV=1.0 h⁻¹
- WHSV=1.4 h⁻¹

T=330 C

- Cs ppm
- % VEGETABLE OIL

- WHSV=0.7 h⁻¹
- WHSV=1.0 h⁻¹
- WHSV=1.4 h⁻¹
Experimental Results for Set 1

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Graph 1: 
- **T=350°C**
- **% VEGETABLE OIL**
- **Cs ppm**
- **WHSV=0.7 h^-1**
- **WHSV=1.0 h^-1**
- **WHSV=1.4 h^-1**

Graph 2: 
- **T=350°C**
- **HCON_tot NL/H2/Kg_feed**
- **% VEGETABLE OIL**
- **WHSV=0.7 h^-1**
- **WHSV=1.0 h^-1**
- **WHSV=1.4 h^-1**

Graph 3: 
- **activity a**
- **time (days)**

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Experimental Reactor Model I

**Kinetic Equations**

- **Hydrodesulphurization**

  \[
  R_{HDS} = k_{HDS} \cdot C_S^n \quad \Rightarrow \quad R_{HDS} = k_{HDS_0} \cdot e^{-\frac{E_{HDS}}{R \cdot T}} \cdot C_S^n
  \]

- **Gasoil Hydrogen Consumption**

  \[
  R_{HCON\_GO} = k_{HCON\_GO_0} \cdot e^{-\frac{E_{HCON\_GO}}{R \cdot T}} \cdot \left(\frac{CHT_{GO} - HCON_{GO}}{ppms \cdot Kg_{feed}} \right)^2
  \]
VO Hydrogen Consumption

\[
R_{HDO} = k_{HDO_0} \cdot e^{\frac{-E_{HDO}}{R \cdot T}} \cdot (C_{VO})_{INITIAL} \cdot (1 - X) \cdot \left(\frac{Kg_{VO}}{hr \cdot Kg_{cat}}\right)
\]

\[
R_{HCON_{VO}} = R_{HDO} \cdot CHT_{VO}
\]

\[
x(z) = \frac{HCON_{VO}(z)}{CHT_{VO}}
\]

\[
R_{HCON_{VO}} = k_{HDO_0} \cdot e^{\frac{-E_{HDO}}{R \cdot T}} \cdot (C_{VO})_{INITIAL} \cdot \left(1 - \frac{HCON_{VO}}{CHT_{VO}}\right) \cdot CHT_{VO} \cdot \left(\frac{NL_{H2_{VO}}}{hr \cdot Kg_{cat}}\right)
\]
Experimental Reactor Model III

Mass Balances

- **Organic Sulphur**
  \[
  \frac{dC_S}{dt} = \frac{-R_{HDS}}{M_{FEED}} \cdot dm_{cat}
  \]  
  (ppms)

- **Hydrogen Sulphide**
  \[
  \frac{dN_{VH2S}}{dt} = \frac{22.4}{MB\cdot S\cdot 10^{-6}} \cdot R_{HDS} \cdot dm_{cat}
  \]  
  \(\frac{NL_{H2S}}{hr}\)

- **Hydrogen**
  \[
  \frac{dN_{VH2}}{dt} = -(R_{HCON\_GO} + R_{HCON\_VO}) \cdot dm_{cat}
  \]  
  \(\frac{NL_{H2}}{hr}\)
Experimental Reactor Model IV

- **Gasoil Hydrogen Consumption**

\[
d_{HCON_{GO}} = \frac{R_{HCON\_GO}}{M_{feed} \cdot \left(1 - \left(C_{VO}\right)_0\right)} \cdot dm \text{ cat}
\]

- **Vegetable Oil Hydrogen Consumption**

\[
d_{HCON_{VO}} = \frac{R_{HCON\_VO}}{M_{feed} \cdot \left(C_{VO}\right)_0} \cdot dm \text{ cat}
\]

- **Total Hydrogen Consumption**

\[
HCON_{TOT} = HCON_{GO} \cdot \left[1 - \left(C_{VO}\right)_0\right] + HCON_{VO} \cdot \left(C_{VO}\right)_0
\]

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Estimation of HDS Kinetic Constants

\[ Y = k_{\text{HDS}} \cdot X \]

\[ Y = \frac{1}{a} \left[ \left( \frac{C_{\text{in}}}{C_{\text{out}}} \right)^{1-n} - 1 \right] \]

\[ X = \frac{1}{\text{WHSV}} \]

**T=350 C**

- R² = 0.9246
- R² = 0.9464
- R² = 0.9583
- R² = 0.984

**T=330 C**

- R² = 0.9775
- R² = 0.9993
- R² = 0.9251
- R² = 0.9485

**T=310 C**

- R² = 0.9282
- R² = 0.9993
- R² = 0.9992
- R² = 0.8392

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HDS Kinetic Constants

\[ n = 1.6 \]

\[ k_{\text{HDS}} \]

\[ -\ln(K_{\text{HDS}}) \]

\[ \frac{1}{T} \]

<table>
<thead>
<tr>
<th>Oil Content</th>
<th>( k_{\text{HDS} , 0} )</th>
<th>( E_{\text{HDS}} )</th>
</tr>
</thead>
<tbody>
<tr>
<td>wt % Palmoil</td>
<td>( \text{ppms}^{(1-n)} \times \frac{K_{\text{feed}}}{K_{\text{cat}}} )</td>
<td>(KJ/mol)</td>
</tr>
<tr>
<td>0.0</td>
<td>8.942E+09</td>
<td>128</td>
</tr>
<tr>
<td>2.5</td>
<td>9.128E+08</td>
<td>120</td>
</tr>
<tr>
<td>5.0</td>
<td>4.112E+08</td>
<td>105</td>
</tr>
<tr>
<td>10.0</td>
<td>2.368E+07</td>
<td>102</td>
</tr>
</tbody>
</table>

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Estimation of Hydrogen Consumption Kinetic Constants

\[ Y = k \cdot \frac{HCON_{GO}}{X} \]

\[ HCON_{GO} = \frac{(HCON_{TOT})_{exp} - CHT_{VO}(C_{VO})_0}{1 - (C_{VO})_0} \]

\[ X = \frac{1}{WHSV \cdot \left[ 1 - (C_{VO})_0 \right]} \]

\[ Y = \frac{1}{a} \left( \frac{1}{CHT_{GO} - HCON_{GO}} - \frac{1}{CHT_{GO}} \right) \]

Graphs showing the relationship between temperature, hydrogen consumption, and conversion rate at different temperatures with various conversion rates.
Hydrogen Consumption Kinetic Constants

% VEGETABLE OIL

K_{HCON_{GO}} = \frac{Kg_{GO}^2}{NLH_2/ hr/Kg_{cat}}

E_{HCON_{GO}} (KJ/mol)

14.5

34.9

CHT_{VO} = 210

NLH_2/ Kg_{VO}

40 % Selectivity towards Hydrogenation/Hydration

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Experimental Conditions

- Liquid Feed: Gasoil (0.5% S) neat or in mixtures with Palm Oil (5.0 and 10.0 % wt):
  - Catalyst: 40 g CoMo/γ-Al2O3 (Albemarle KF 757), diluted with 50 g SiC.
  - Reaction Temperature: 330 – 350 – 365 °C
  - Reactor Pressure: 33 bar
  - Liquid flow rate: 56, 40 and 34 g/h, WHSV: 1.4, 1.0 and 0.85 g<sub>oil</sub>/g<sub>cat</sub>/h
  - Gas flow rate: 20 Nl<sub>H2</sub>/h

Analysis

- Sulphur content in product oil, Exit gas composition
Experimental Results for Set 2

T=350 °C

T=330 °C

T=365 °C

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\[ Y = k_{HDS} \cdot X \]

\[ X = \frac{1}{\text{WHSV}} \]

\[ Y = \frac{1}{a} \left[ \frac{(C_{\text{in}})^{1-n} - (C_{\text{out}})^{1-n}}{(n-1)} \right] \]

- **T=350°C**
  - \( R^2 = 0.9983 \)
  - \( R^2 = 0.9999 \)
  - \( R^2 = 0.9027 \)

- **T=365°C**
  - \( R^2 = 0.9804 \)
  - \( R^2 = 0.9932 \)
  - \( R^2 = 0.9995 \)

- **T=330°C**
  - \( R^2 = 0.913 \)
  - \( R^2 = 0.9901 \)
  - \( R^2 = 0.9601 \)

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Estimation of HDS kinetic constants / Set II

UGent/FCh13/3L
**HDS kinetic constants / Set II**

**n=1,55**

![Graph showing % VEGETABLE OIL vs. \( k_{HDS} \) (ppms\(^{(1-n)}\) Kg\(_{feed}\)/h/Kg\(_{cat}\)).](Image)

**Table:**

<table>
<thead>
<tr>
<th>CONTENT</th>
<th>( k_{HDS} )</th>
<th>( E_{HDS} )</th>
</tr>
</thead>
<tbody>
<tr>
<td>% wt Palm oil</td>
<td>( \text{(ppms}^{(1-n)} \ast \text{Kg}<em>{\text{feed}})/\text{h/Kg}</em>{\text{cat}}) )</td>
<td>(KJ/mol)</td>
</tr>
<tr>
<td>0.0</td>
<td>5.20E+11</td>
<td>145</td>
</tr>
<tr>
<td>5.0</td>
<td>8.19E+07</td>
<td>104</td>
</tr>
<tr>
<td>10.0</td>
<td>2.59E+07</td>
<td>98</td>
</tr>
</tbody>
</table>

**Graphs:**

- **Graph 1:** \(-\text{LN (K}_{HDS}\) vs. \(1/T\))
  - \(R^2 = 0.9999\)
  - \(R^2 = 0.9983\)
  - \(R^2 = 0.9821\)

- **Graph 2:** \(Cs_{\text{exp}}\) (ppmwS) vs. \(Cs_{\text{calc}}\) (ppmwS)
  - 0.0 %
  - 5.0 %
  - 10.0 %

**Notes:**

- 1 March 2013
- UGent/FCh13/3L
Gas analysis

<table>
<thead>
<tr>
<th>Gas</th>
<th>330°C/1.4 h⁻¹</th>
<th>350°C/1.4 h⁻¹</th>
<th>350°C/1.0 h⁻¹</th>
</tr>
</thead>
<tbody>
<tr>
<td>H₂</td>
<td>96.73%v/v</td>
<td>96.83%v/v</td>
<td>97.62%v/v</td>
</tr>
<tr>
<td>H₂S</td>
<td>0.94%v/v</td>
<td>1.12%v/v</td>
<td>0.59%v/v</td>
</tr>
<tr>
<td>CO</td>
<td>0.69%v/v</td>
<td>0.64%v/v</td>
<td>0.36%v/v</td>
</tr>
<tr>
<td>CO₂</td>
<td>0.69%v/v</td>
<td>0.44%v/v</td>
<td>0.54%v/v</td>
</tr>
<tr>
<td>CH₄</td>
<td>0.51%v/v</td>
<td>0.55%v/v</td>
<td>0.51%v/v</td>
</tr>
<tr>
<td>C₃H₈</td>
<td>0.44%v/v</td>
<td>0.42%v/v</td>
<td>0.39%v/v</td>
</tr>
</tbody>
</table>

Vegetable Oil in Feed : 10 wt %
Algorithm of calculations

1. Determination of gaseous molar fractions, \( y_i \) from the gas analysis.
2. By fitting this molar fractions, the ones in the separator’s inlet, \( z_i \) were calculated using a cubic equation of state.
3. These molar fractions are the same at the reactor’s outlet.
4. Molar flows of the gaseous products calculated from the molar fractions are used for further calculations.

- The sum of CO, CO\(_2\) and CH\(_4\) molar flows at Reactor Outlet give the amount of ester bonds following the paths of decarboxylation and decarbonylation.
  - Combining with V.O. structure calculate catalyst selectivity.

Dehydration pathway is preferred by 38% of the total ester bonds and the remaining (62%) follow the other two paths.
Operating Parameters of the Industrial Unit

- Liquid Inlet Flow Rate = 120 - 170 tn/hr
- Gas Inlet Flow Rate (Recycle + Make-up) = 48000 - 55000 Nm$^3$/hr
- Quench 1 Flow Rate = 3000 Nm$^3$/hr (75 % vol/vol Hydrogen)
- Quench 2 Flow Rate = 2000 Nm$^3$/hr (75 % vol/vol Hydrogen)
- $T_{\text{Inlet}} = 280 - 345 \, ^{\circ}\text{C}$
- $P = 34 - 39 \, \text{bar}$
- Catalyst Mass $M_{\text{cat}} = 120 \, \text{tn}$
Industrial Reactor Model

Mass Balances

- Organic Sulphur
- Hydrogen Sulphide
- Vegetable Oil Hydrogen Consumption
- Gasoil Hydrogen Consumption

Energy Balance

\[ dT = \frac{R_{\text{HCON\_GO}} \cdot \Delta H_{R\_GO}(T) + R_{\text{HCON\_VO}} \cdot \Delta H_{R\_VO}(T)}{\sum F_i \cdot C_p_i(T)} \cdot dm_{\text{cat}} \]

Unsaturated C-C bonds: \( \Delta H_{R\_GO} = 1.4 \text{ KJ/NL}_H2 \)

Ester bonds deoxygenation \( \Delta H_{R\_VO} = 2.5 \text{ KJ/NL}_H2 \)
Hydrogen Consumption and Temperature Profiles

Low Sulphur Gasoil

Cs_in = 5000 ppm

T=350 C, WHSV=1.4 h⁻¹, 0 % V.O

T=350 C, WHSV=1.4 h⁻¹, 5% V.O

Results

UGent/FCh13/3L
Organic Sulphur and Hydrogen Flow Rate Profiles

T=350 C, WHSV=1.4 h⁻¹, 0 % V.O

T=350 C, WHSV=1.4 h⁻¹, 5% V.O
Effects of the Operating Parameters

HEAVY FEED (C_{\text{sin}}=8800 \text{ ppm})

LIGHT FEED KK3 (C_{\text{sin}}=5000 \text{ ppm})

Results

UGent/FCh13/3L

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Effects of the V.O. Content / Light Feed

HCON + 2,0-2,3 NLH₂/Kg_feed per 1 % V.O.

T_out + 1,3 °C per 1 % V.O.
**Effects of the Inlet Temperature/ Light Feed**

HCON + 2,6-3,1 NLH2/Kg_feed per ΔT_in=10 °C

T_out + 1,0 °C per ΔT_in=1 °C

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Effects of the Space Velocity / Light Feed

Results

0.0 % V.O.

10.0 % V.O.

0,5 1,0 1,5 2,0 2,5

WHSV (h⁻¹)

0,5 1,0 1,5 2,0 2,5

WHSV (h⁻¹)

0.0 % V.O.

10.0 % V.O.

0,5 1,0 1,5 2,0 2,5

WHSV (h⁻¹)

0,5 1,0 1,5 2,0 2,5

WHSV (h⁻¹)

UGent/FCh13/3L
Effects of the Space Velocity / Light Feed II

HCON - 1,5-1,9 NLH2/Kg_feed per ΔWHSV=0,1 h⁻¹
<table>
<thead>
<tr>
<th>Comparison</th>
<th>Biodiesel (FAME)</th>
<th>Green Diesel</th>
</tr>
</thead>
<tbody>
<tr>
<td>% O</td>
<td>11</td>
<td>0</td>
</tr>
<tr>
<td>Density, g/ml</td>
<td>0.88</td>
<td>0.78</td>
</tr>
<tr>
<td>Sulfur Content, ppm</td>
<td>&lt; 10</td>
<td>&lt; 10</td>
</tr>
<tr>
<td>Energy Content MJ/kg</td>
<td>38</td>
<td>44</td>
</tr>
<tr>
<td>% Difference in NOx Emissions</td>
<td>+10</td>
<td>0 – (-10)</td>
</tr>
<tr>
<td>Neat</td>
<td>Liquid</td>
<td>Solid</td>
</tr>
<tr>
<td>Distillation 10 – 90 %</td>
<td>340 - 355</td>
<td>265 - 320</td>
</tr>
<tr>
<td>Cetane Number</td>
<td>50</td>
<td>80 - 90</td>
</tr>
</tbody>
</table>
Conclusions


- At conventional HDT conditions total V.O. conversion to Green Diesel is achieved.

- Deactivation of Co-hydroprocessing is similar to that during Gasoil HDT.

- The increase of the V.O. content up to 5% in the reactor feed results in a decrease of HDS reaction rates, but for a further increase up to 10% the desulphurization rates remained unchanged.

- The selectivity of the CoMo/Al2O3 catalyst is 38% for hehydration reactions and 62% for hydro-decarbonylation/decarboxylation.

- The increase of the V.O. content in the feed results in increase of the produced heat in the reactor and the temperature difference $T_{OUT} - T_{IN}$. 

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• Collaborators:
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  - Sebos Ioannis  Ph.D. Student
  - Templis Chrysovalantis  Ph. D. Student
  - Vonortas Andreas Ph. D. Student

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  - GF Energy

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