Lecture Series **Heterogeneous Catalysis**

Fritz-Haber-Institute (Max Planck Society)
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**Micro-Structured Reactors and Catalysis**

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Principles of Micro-Structured Reactors: 
Application in Catalysis & Process Engineering

1 : $10^{-12}$

Institut für Mikrotechnik
Mainz
Subjects of Presentation

1. Principles of Micro-Structured Reactors (MSR)

2. MSR as Tool in Catalysis
   2.1 Provision of kinetic data
   2.2 (in-situ) Characterization of catalytic materials
   2.3 Application of MSR in industrial catalysis
      - Hydrogen-driven fuel cells
      - Propene oxidation by $\text{H}_2\text{O}_2$ vapor
## Scales of chemical reactors

<table>
<thead>
<tr>
<th></th>
<th>Industry</th>
<th>Laboratory</th>
<th>Microsystem</th>
</tr>
</thead>
<tbody>
<tr>
<td><strong>Volume</strong></td>
<td>30 m³</td>
<td>10⁻³ m³</td>
<td>3 × 10⁻¹¹ m³</td>
</tr>
<tr>
<td><strong>Scale-down</strong></td>
<td>1</td>
<td>1:3 × 10⁻⁵</td>
<td>1: 10⁻¹²</td>
</tr>
<tr>
<td><strong>Diameter</strong></td>
<td>2 m</td>
<td>2 cm</td>
<td>20 μm</td>
</tr>
<tr>
<td><strong>Surface/Volume</strong></td>
<td>2 m²/m³</td>
<td>200 m²/m³</td>
<td>200'000 m²/m³</td>
</tr>
</tbody>
</table>
Laboratory

Pilot-plant

Production

scale-up

« scale up »
Micro-structured multichannel reactors

Volume: $V_R = 5 \text{ cm}^3$
Pressure drop: $\Delta p = 1 \text{ bar}$

length: $L = 5.$ cm
diameter: $d = 100.$ $\mu$m
Number of channels: $N = 12,740.$
Specific surface: $a = 40,000.$ $\text{m}^2/\text{m}^3$
Flow velocity: $u = 0.63 \text{ m/s (H}_2\text{O)}$
$\text{35. m/s (air)}$
Mass flow: $Q_m = 225.$ $\text{kg/h (H}_2\text{O)}$
$15.$ $\text{kg/h (air)}$
Typical time and length scale of chemical reactors

- Micro-structured apparatus
- Milli-structured apparatus
- Multi-tube reactor
- Tube reactor
- Stirred tank reactor

Length scale, [m]
- 1 μm
- 10 μm
- 100 μm
- 1 mm
- 10 mm
- 100 mm
- 1 m
- 10 m

Mean residence time, [s]
- $10^{-2}$
- $10^{-1}$
- $10^{0}$
- $10^{1}$
- $10^{2}$
- $10^{3}$
- $10^{4}$
Mass transfer in heterogeneous catalysis

- Product
- Feed
- Catalyst
- Feed gas
- Coolant in
- Coolant out
- Tubes filled with catalyst

[Images of catalysts and process flow diagrams]
Scales of Heterogeneous Solid Catalysts

(MSR: $5 \cdot 10^{-3} - 5 \cdot 10^{-2}$ cm)
**Process Steps in Heterogeneous Catalytic Reactions**

Steps during the course of the reaction:
1. External diffusion
2. **Internal diffusion**
3. Adsorption on the active sites
4. Surface reaction forming the products
5. Desorption of the products
6. **Internal diffusion**
7. External diffusion

**Diagram:**
- **Substrate A** to **Product P**
- **Boundary layer**
- **Fixed bed reactor**

*Agar, 2003*
Internal mass transfer

- Internal mass transfer resistance can be reduced by
  - small particles
  - thin catalytic layers
Internal mass transfer

- Internal mass transfer resistance can be reduced by
  - small particles
  - thin catalytic layers

Internal mass transfer can be neglected for catalytic layers $\delta < 20 \mu m$

$\phi = \delta_{cat} \sqrt{\frac{r_V}{D_{eff} \cdot c_{1,s}}}$; \hspace{1cm} $\psi = \frac{\delta_{cat}^2 \cdot r_{V,eff}}{D_{eff} \cdot c_{1,s}}$

Thiele-modulus \hspace{1cm} Weisz-modulus

Internal mass transfer can be neglected, if

$$\psi = \frac{\delta_{cat}^2 \cdot r_{V,eff}}{D_{eff} \cdot c_{1,s}} < \begin{cases} 0.7; \quad n=0 \\ 0.07; \quad n=1 \\ 0.03; \quad n=2 \end{cases}$$
Transport simultaneous with reaction: internal mass transfer within a spherical particle

The concentration profile established in the particle depends on the ratio between the characteristic diffusion time $t_D$ and the characteristic reaction time $t_r$

$$\frac{t_D}{t_r} = \frac{L^2}{D_e} k_r$$

First order reaction

$$\phi = L \sqrt{\frac{k_r}{D_{ef}}}$$

$L$ is a characteristic length (the radius for the sphere particle), and $D_e$ is effective diffusivity of a species A in the particle
Temperature control in catalytic wall reactors

- Heat production near the channel wall;
- Main resistance to heat transfer in the catalytic layer;

Near isothermal conditions, if

\[
\frac{\left|\Delta H_R\right| \cdot r_{V,\text{eff}} \cdot \delta_{\text{cat}}^2}{\lambda_{\text{cat}} \cdot T_w} < 0.15 \frac{RT_w}{E_a}
\]
Properties of Micro-Structured Reactors

- High surface area per reactor volume (beneficial for heat exchange)
- Short contact (reaction) time
- Internal (and external) mass transport of reactants can be neglected
- Negligible resistance to heat transport from the catalyst layer to the wall
- Low pressure drop as compared to a packed bed
- Suppression of danger of explosion
Use of MSR as a Tool in Catalysis

Determination of isothermal kinetic data

A. Oxidative dehydrogenation of propane on $\text{VO}_x$ catalysts
Oxidative Dehydrogenation of Propane (ODP)

\[ \Delta_{R}H_{873K} = -118 \text{ kJ mol}^{-1} \]

\[ \text{C}_3\text{H}_8 + \text{O}_2 \rightarrow \text{C}_3\text{H}_6 + \text{O}_2 \rightarrow \text{CO}_x \]

\[ \Delta_{R}H_{873K} = -1195 \text{ to } -2045 \text{ kJ mol}^{-1} \]

conventional fixed-bed reactor

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Assembly of the micro-channel reactor

Micro-channel reactor (IMM)  
housing plate (stainless steel)

Single plate with deposited catalytic material (VO$_x$/Al$_2$O$_3$)

Channel dimension: 20 mm x 500 µm x 100 µm

10 micro-structured plates (titanium)
CFD simulation of temperature profiles for ODP in a single micro-channel (T = 773 K)

\[ C_3H_8/O_2/Ne = 0.5/0.25/0.25 \]

- **Catalysis**
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Isothermal condition within the channels when a reactor material is used with high heat conductivity and thick walls

Heat conductivity of the wall = 15 W/m/s
Wall thickness: 125/150 µm
Comparison of a micro-structured and a packed-bed reactor

- Micro-structured wall reactor
- Packed-bed reactor

- **T** - gradient
- **C\textsubscript{3}H\textsubscript{8}** conversion
- **O\textsubscript{2}** conversion

\[ T_{\text{max}} - T_{\text{inlet}} / \text{K} \]

- P_{\text{C3H8}} = 51.2 \text{ kPa}, \ P_{\text{O2}} = 25.6 \text{ kPa}, \ p_{\text{ges}} = 101.2 \text{ kPa}

- **packed-bed reactor:** severe temperature gradients
- **micro-structured wall reactor:** almost no temperature gradients
Re-oxidation

2 [ ] + O₂ → 2 [O]

[O] – lattice oxygen    [ ] – reduced VOₓ-site
## Kinetic parameters

*(formation rates of C\textsubscript{3}H\textsubscript{6} and CO\textsubscript{x} \propto [O]*)

<table>
<thead>
<tr>
<th>Reaction</th>
<th>Rate Constant ( k )</th>
<th>( k_{733 , K} ) ( \text{mol kg}^{-1} \text{s}^{-1} )</th>
<th>( E_A ) / ( \text{kJ mol}^{-1} )</th>
</tr>
</thead>
<tbody>
<tr>
<td>( \text{C}_3\text{H}_8 \rightarrow \text{C}_3\text{H}_6 )</td>
<td>( k_1 ) ( p_{\text{C}_3\text{H}_8} [\text{O}] )</td>
<td>2.9 ( \cdot ) 10\textsuperscript{-6}</td>
<td>121</td>
</tr>
<tr>
<td>( \text{C}_3\text{H}_6 \rightarrow \text{CO} )</td>
<td>( k_2 ) ( p_{\text{C}_3\text{H}_6} [\text{O}] )</td>
<td>9.7 ( \cdot ) 10\textsuperscript{-6}</td>
<td>102</td>
</tr>
<tr>
<td>( \text{C}_3\text{H}_6 \rightarrow \text{CO}_2 )</td>
<td>( k_3 ) ( p_{\text{C}_3\text{H}_6} [\text{O}] )</td>
<td>8.3 ( \cdot ) 10\textsuperscript{-6}</td>
<td>98</td>
</tr>
<tr>
<td>( \text{C}_3\text{H}_8 \rightarrow \text{CO} )</td>
<td>( k_4 ) ( p_{\text{C}_3\text{H}_8} [\text{O}] )</td>
<td>3.7 ( \cdot ) 10\textsuperscript{-7}</td>
<td>155</td>
</tr>
<tr>
<td>( \text{C}_3\text{H}_8 \rightarrow \text{CO}_2 )</td>
<td>( k_5 ) ( p_{\text{C}_3\text{H}_8} [\text{O}] )</td>
<td>3.2 ( \cdot ) 10\textsuperscript{-7}</td>
<td>126</td>
</tr>
<tr>
<td>( \text{O}_2 \rightarrow [\text{O}] )</td>
<td>( k_6 ) ( p_{\text{O}_2} [\text{O}] )</td>
<td>3.7 ( \cdot ) 10\textsuperscript{-6}</td>
<td>87</td>
</tr>
</tbody>
</table>
Parity Plot (rates of formation of $\text{C}_3\text{H}_6$ and $\text{CO}_x \propto [\text{O}]$)

Open symbols:

- $T = 693-796 \text{ K}$
- $\text{C}_3\text{H}_8/\text{O}_2/\text{Ne}$
  - $0.3/0.15/0.55$
  - $0.50/0.25/0.25$
  - $0.15/0.15/0.70$
# Simulation of the lab-scale fixed-bed reactor

<table>
<thead>
<tr>
<th>$T_{\text{Inlet}}$ / K</th>
<th>$\Delta T$ / K</th>
<th>$X(C_3H_8)$ / %</th>
<th>$X(O_2)$ /%</th>
<th>$S(C_3H_6)$ /%</th>
<th>$S(CO)$ /%</th>
<th>$S(CO_2)$ /%</th>
</tr>
</thead>
<tbody>
<tr>
<td>693</td>
<td>1</td>
<td>3.0</td>
<td>8.5</td>
<td>77</td>
<td>13</td>
<td>10</td>
</tr>
<tr>
<td></td>
<td>3</td>
<td>2.9</td>
<td>9.0</td>
<td>76</td>
<td>12</td>
<td>12</td>
</tr>
<tr>
<td>773</td>
<td>20</td>
<td>17.1</td>
<td>82.4</td>
<td>47</td>
<td>35</td>
<td>18</td>
</tr>
<tr>
<td></td>
<td>28</td>
<td>14.8</td>
<td>58.3</td>
<td>57</td>
<td>25</td>
<td>17</td>
</tr>
</tbody>
</table>

*blue – simulated values*
Use of MSR as a Tool in Catalysis

Determination of isothermal kinetic data

B. Ammonia oxidation on a thin Pt foil
Ammonia oxidation on Pt
Temperature control during Reaction

Temperature-programmed reaction over:

Pt gauze in tubular reactor

$\text{NH}_3 : \text{O}_2 = 3:3$
red. Pt gauze

$400 \text{ ml/min}$
$101 \text{ kPa}$
$3\% \text{ NH}_3$
$3\% \text{ O}_2$

Ignition

$T_{\text{sandbed}} / \degree \text{C}$

NH$_3$ conversion / vol% 0 20 40 60 80 100

Pt foil in the micro-structured reactor

$\text{NH}_3 : \text{O}_2 = 3:4.5$
red Pt foil

$250 \text{ ml/min}$
$101 \text{ kPa}$
$3\% \text{ NH}_3$
$4.5\% \text{ O}_2$

NH$_3$ conversion / vol% 0 10 20 30 40 50

$T_{\text{catalyst}} / \degree \text{C}$

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poor heat transfer leads to ignition = loss of temperature control

micro reactor provides temperature control up to 700$\degree \text{C}$
Surface reconstruction of thin-foil Pt catalyst

- Assembled reactor
- Exploded view
- Cut view A-A

- Simple flat catalyst in optimal contact with the wall
- Fast catalyst exchange and protected thermocouple
- Short contact time (for fast reactions)

Operates without gaskets up to 130 kPa!
**Measured influence of temperature on rates of product formation**

\[ r_i = f(T) \]

N\textsubscript{2}, N\textsubscript{2}O, and NO are formed as products.

- Product formation increases with temperature.
- NO formation requires T > 330°C.

**Experimental Conditions:**
- NH\textsubscript{3} : O\textsubscript{2} = 1:3
- 4xN\textsubscript{2}O
- T / °C: 250, 300, 350, 400, 450
- 400 ml/min
- 101 kPa
- X<11%
- TOS>14h
Measured influence of adding products NO and N$_2$O on rates of product formation

(NH$_3$+O$_2$) + NO

\[ r_i = f(p_{NO}) \]

\[ \begin{array}{c}
\text{400 ml/min} \\
\text{101 kPa} \\
\text{X<11%} \\
\text{TOS>14h} \\
\text{3 kPa NH}_3 \\
\text{3 kPa O}_2
\end{array} \]

(NO reduces rate of nitrogen formation, but increases it slightly for N$_2$O formation)

(NH$_3$+O$_2$) + N$_2$O

\[ r_i = f(p_{N2O}) \]

\[ \begin{array}{c}
\text{400 ml/min} \\
\text{101 kPa} \\
\text{X<11%} \\
\text{TOS>14h} \\
\text{3 kPa NH}_3 \\
\text{3 kPa O}_2
\end{array} \]

N$_2$O has hardly any influence on rate of product formation
Ammonia oxidation on thin-film Pt catalyst

Temperature-dependent surface reconstruction

SEM after reaction at:

- 330°C
- 374°C
- ~700°C

- 5 µm
- 1 µm
- 2 µm

Temperature-dependent surface faceting

Different Pt morphology at low- and high-T <=> different transport mechanism?

T-controlled reaction below 500°C made it possible to study faceting & kinetics

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Building models:
Set of Reactions and Rate equations

simplified reaction network

NH₃ adsorption
NH₃ + s \rightarrow NH₃-s
NH₃-s \rightarrow NH₃ + s

O₂ adsorption
O₂ + 2s \rightarrow 2 O-s
2 O-s \rightarrow O₂ + 2s

NH₃ activation
NH₃-s + 3/2 O-s \rightarrow N-s + 3/2 H₂O + 3/2 s
r = k_i (NH₃-s)(O-s)

H₂O formation

N₂ formation reactions
2 N-s \rightarrow N₂ + 2 s

NO formation
N-s + O-s \rightarrow NO-s + s

NO desorption adsorption
NO-s \rightarrow NO + s
NO + s \rightarrow NO-s

N₂O formation reactions
NO-s + N-s \rightarrow N₂O + 2s
Mechanistic indications from kinetic fitting

best-fitting model

286-385°C

\[ p_i < 6 \text{ kPa} \]

- re-adsorption of NO must be considered
- decomposition of NO is not significant
- adsorption sites and reactions are consistent with literature on Pt(111)

A kinetic model on pc Pt describes the formation of \( \text{N}_2 \) and \( \text{N}_2\text{O} \) and NO

The Pt surface is reconstructed already at mild conditions (374°C); this knowledge is essential for evaluation of kinetic data!
Conclusions

Micro-structured reactors are suited for obtaining isothermal kinetic data

A.) for simulation of catalytic fixed-bed reactor performance and

B.) for deriving indications on possible reaction mechanisms
Use of MSR as a Tool in Catalysis

- Surface restructuring in ammonia oxidation on a Pt foil as catalyst at defined $T$ (remember previous section)
- Identification of surface oxides of a Mo/V/W-$O_x$ on silica catalyst by Laser-Raman spectroscopy
FHI

In-situ Laser Raman Catalyst Characterization
- C₃H₆ oxidation by catalyst on carrier plate -

assembled transparent quartz reactor

exploded view

cut view

140 mm

300 µm

125 µm

T

gas

screws / clamp

heaters

sand blasted

+ cheap simple support plates

+ defined catalyst film

+ fast catalyst exchange (high contact time)

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Catalysis
Laser-Raman Setup

- Laser (633 nm)
- Slit
- Confocal hole
- Notch filter
- CCD (243K)
- Laser (633 nm)
- Confocal Raman spectrometer
- Catalyst plate
- Heaters
- MFC
- Analysis
- PTR-MS
- O₂
- C₃H₆
- He

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Propene oxidation / Experimental

**Reactor A + Raman**
- Reactor A
- Raman
- Flow: 1 ml/min
- \( \tau \): < 500 ms
- \( \text{Re} \): < 1
- \( T \): < 400°C
- \( p \): 101 kPa
- \( p_i \): < 20 kPa

**Catalyst**
- Catalyst preparation:
  - Mo-V-W + gelatine + water
  - Oxidized Si wafer
  - *solution (Mo, W, V)
  - *spray drying (powder)
  - *add water-gelatine
  - *spin-coat at 37°C, dry
  - *calcine 3h@300°C in air

**Catalyst**
- Mo\(_{0.68}\)V\(_{0.23}\)W\(_{0.09}\)O\(_x\)
- SiO\(_2\)
- Si(100)
- Mo-V-W-Oxid
- 200 nm

**Reactants**
- \( C_3H_6 + O_2 \)
- Acrolein
- Acrylic acid
- Acetaldehyde
- Acetic acid
- CO
- CO\(_2\)
**Propene oxidation** (in-situ Laser-Raman cat. char.)

**T-programmed reaction**

- 1 ml/min
- 10% C₃H₆
- 20% O₂
- 70% He
- X < 5%

![Reaction Graph]

**in-situ Raman**

- *MoO₃
- **Mo₅O₁₄
- 400°C
- 380°C
- 300°C

- * active catalyst / all products
- * activation + change of Raman = structure vs. activity in parallel

**Note:**
- A. Brückner / ACA

**Catalysis**
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Use of MSR as a Tool in Applied Catalysis

A. Hydrogen-driven fuel cells
B. Partial oxidation of propene by H$_2$O in the vapor phase
Aspects of micro-structured reactor applications for mobile fuel cells

- choice of hydrogen carrier for hydrogen-driven fuel cells
- for maximizing hydrogen selectivity in steam reforming of methanol, low temperatures are required for thermodynamic reasons
- coating of ceramic walls with catalytic material
Fueling fuel cells for mobile applications

Hydrogen storage is not ideal for portable devices.

A liquid fuel (hydrogen-carrier) is stored and reformed to produce on-demand hydrogen for the fuel cell.

Which fuel should be used?

Prototype reformer and PEMFC for a laptop from Casio, about 20 hr of operation.


University of New Mexico
Energy densities for various “hydrogen-carriers”

Volumes and weights of different fuels equivalent to 150Wh of stored energy*

<table>
<thead>
<tr>
<th>Fuel</th>
<th>Volume</th>
<th>Weight</th>
</tr>
</thead>
<tbody>
<tr>
<td>Li-ion Battery</td>
<td>600 ml</td>
<td>1 kg</td>
</tr>
<tr>
<td>H₂ Compressed (200 bar)</td>
<td>340 ml</td>
<td>920 g</td>
</tr>
<tr>
<td>Liquid Hydrogen</td>
<td>95 ml</td>
<td></td>
</tr>
<tr>
<td>Metal Hydride</td>
<td>95 ml</td>
<td></td>
</tr>
<tr>
<td>Methanol</td>
<td>33 ml</td>
<td>26 g</td>
</tr>
<tr>
<td>Diesel</td>
<td>13 ml</td>
<td>13 g</td>
</tr>
</tbody>
</table>

Why Methanol

- Low reforming temperature (220-250 °C)
- Lower CO production compared to other fuels.
- High H₂ content in product stream
- Zero sulfur

Steam reforming of methanol

\[ CH_3OH + H_2O \rightarrow 3H_2 + CO_2 \]

\[ CO_2 + H_2 \leftrightarrow CO + H_2O \]

CO production must be minimized because it is a poison to the anode catalyst.

For a 20W fuel cell, assuming a liquid flow of 15ml/hr and 98% MeOH conversion approximately 4.8g of catalyst are needed.
Methanol Conversion for Different Fixed-Bed Reactor Dimensions

Steam/Methanol = 1.1, T = 230°C, P = 640 Torr

Packed Bed ID
- 4.1 mm
- 1.75 mm
- 1.0 mm

Methanol conversion vs. W/F (kg s/gmol)

University of New Mexico
Packed-bed reactors are far from isothermal
(Enthalpy of reforming reaction: ca. -41 kJ/mol)

Reactor: 1mm ID
$U_0 = 0.26 \text{ m/s}$
$W/F = 16 \text{ kg s/gmol}$
Methanol Conversion = 0.86

$\Delta T$ is 20K for such a small-diameter reactor


University of New Mexico
Solution:
micro channel reactors
(ca. 500 to 100 μm)

Coating of micro-structured ceramic tubes
Gas-Assisted Fluid Displacement

First, the fused-silica capillary is cleaned by etching, then rinsed with water and subsequently with ethanol; finally it is dried in air.

Afterwards, the capillary is filled with the coating fluid.

Purging the tube with air causes a thin film of liquid to be left behind.

\[ m = \frac{(coated\ mass) - (empty\ mass)}{(full\ mass) - (empty\ mass)} \]
Wall coated reactor represents superior geometry

530μm capillary showing a dried 15μm thick Cu/ZnO/Al$_2$O$_3$ catalyst coating.

Wall coated reactors are better than micro-packed bed reactors for catalyst incorporation:

- Inherently low pressure drop.
- Short diffusion/conduction lengths lead to isothermal operation.
- More stable and reliable than packed bed for long-term portable use.
Radial Uniformity

~5 microns
Axial Uniformity

Axial uniformity is determined by comparing points on the ends and in the middle of the coated capillary.
Multi-channeled structure fabrication

Fused-silica capillaries

Capillaries bundled and set with high T epoxy.

Ends cut with ceramic saw.

Capillary bundle inserted into s.s. housing and sealed with epoxy.

¼” OD stainless steel tube
Multi-Channeled Structures

Transmitted-light optical microscope image of an un-coated M.R. with about 60 250μm capillaries.

The channels are regularly spaced and the epoxy has filled in all of the voids between the tubes creating an air-tight seal.
Micro-structured fuel-cell processor
Complete Microstructured Fuel Processor for Methanol Steam Reforming – (Running for 5 Days at Hannover Fair 2005)

Methanol + Water

Micro-structured Evaporator (250°C)

Catalyst: Industry

Reformer (270°C)

≈ 1% CO

Reactor Size 7 x 2 x 1 cm³

400 ml/min H₂ produced

Oxygen

O₂/C = 4

Preferential Oxidation Stage 2 (140°C)

< 100 ppm CO

Preferential Oxidation Stage 1 (140°C)

< 15 ppm CO

Catalyst: Home-made

Fuel Cell (30 W) (PEM)

Catalyst: Home-made

Electrical Consumer

CO₂, Water

Catalytic Burner (200°C)

Catalyst: Pt (DEGUSSA)

Oxygen

O₂/C = 8

Condensation

IR-Sensor

All devices were externally heated

Preferential oxidation of CO (PrOX) in a micro-structured reactor

Specifications:
• 53 microstructured plates
• Volume: 60 cm$^3$
• Mass: 150 g
• 1.5 g Pt/Ru/Al$_2$O$_3$ catalyst
• coating thickness: 50 mm

Experimental conditions:
• simulated reformate gas:
  $H_2/CO_2/H_2O/CO/O_2 = 56/18/10/0.5/0.9 \%$
• coolant gas: nitrogen


IChemE Chemistry Innovation KTN/Impact Award
TU/e award winner for TU/e-IMM development
(Whitehall, London)
Testing rig for the steam-reforming reaction in micro-structured reactors
Demonstration plant reactor
consisting of
micro-structured catalytic modules
for epoxidation of
propene to propene oxide
by vapor-phase hydrogen peroxide

DEMIS Consortium:
Degussa, Uhde/Thyssen-Krupp, TU Chemnitz und Darmstadt,
MPI-Mühlheim
Gas-phase epoxidation of propene by hydrogen peroxide to propene oxide

towards technical scale

© DEMIS® project
Microplant Composed of Various Modules

- Micro Annular Gear Pump
- Micro Heat Exchanger
- Micromixer
- Micro Gas Separator
- Sampling Valve
- Product
- Air
- Toluene
- SO₃
- H₂O
- Backbone Element
Chemie in Mikrostrukturreaktoren
Klaus Jähnisch,* Volker Hessel,* Holger Löwe und Manfred Baerns

Stichwörter:
Flüssigflüssigreaktionen - Gasflüssigreaktionen - Heterogene Katalyse - Mikrostrukturreaktoren - Prozessoptimierung

\[
\begin{align*}
\text{Me} & \quad 10\% \ F_2 \ \text{in N}_2 \\
\text{Me} & \quad \text{Me} \quad \text{Me} \\
& \quad \text{MeCN} \cdot -15^\circ C \\
\end{align*}
\]

\[
\text{Me} + \text{Me} + \text{Me} + \text{Me} \\
\]
The End

References:

M. Matlosz, W. Ehrfeld, J.P. Baselt (Eds.): Microreaction Technology; Springer, Heidelberg 2001


