

# Data Driven Conceptual Process Design for the Hydroformylation of 1-Dodecene in a Thermomorphic Solvent System

Kevin McBride<sup>†</sup> and Kai Sundmacher<sup>\*,†,‡</sup>

<sup>†</sup>Max Planck Institute for Dynamics of Complex Technical Systems, Sandtorstrasse 1, 39106 Magdeburg, Germany

<sup>‡</sup>Process Systems Engineering, Otto-von-Guericke University Magdeburg, Universitätsplatz 2, 39106 Magdeburg, Germany

**S** Supporting Information

**ABSTRACT:** The purpose of this work is to investigate the effect that a solvent mixture has on the process-wide performance of the hydroformylation of 1-dodecene. A thermomorphic solvent mixture consisting of dimethylformamide and decane is used as a tunable solvent that ensures monophasic conditions during the reaction and that is also used to separate the homogeneous catalyst from the product through temperature controlled phase splitting in a decantation unit. So far reactor and process design for this reaction has been primarily based on a specific composition of the solvent mixture and without considering the economic impact of catalyst leaching. In order to reduce the complexity of the conceptual process design problem, linear regression models were used to accurately estimate the partition coefficients of the multicomponent mixture in the decanter. Using limited available experimental data from the literature, correlations were derived for catalyst loss based on the composition of the thermomorphic solvent system. Optimization results show that the catalyst loss has a significant effect on the total annualized cost of the process and that the frequently studied solvent composition should be altered to increase catalyst retention.

## INTRODUCTION

In the past decade, interest in homogeneous catalysis has increased substantially with the recovery or retention of expensive transition metal catalysts being the object of much study. This is especially true in the hydroformylation of long-chain olefins where, due to the high cost of the catalyst complex composed quite often of rhodium and expensive ligands, industrial realization has been problematic.<sup>1</sup> Many different methods have been investigated primarily for the hydroformylation of 1-octene, which is extensively discussed in the review by Cole-Hamilton.<sup>2</sup> One effective method for hydroformylation has been done using supported ionic liquid phase catalysis, where the catalyst is supported in a thin film formed by an ionic liquid.<sup>3</sup> Other attempts at restricting catalyst loss involve tethering the catalyst complex to mesoporous materials, effectively creating semiheterogeneous catalysts.<sup>4</sup> Membrane filtration methods were also effectively developed for the continuous hydroformylation of 1-octene using specially designed polymer catalysts, whose bulky nature reduces the permeability of the catalyst through the membrane.<sup>5</sup> A CO<sub>2</sub>-expanded liquid (CXL) enhanced version of this process has also been implemented and operated economically with reasonably low and stable levels of catalyst loss.<sup>6</sup> A similar principle has been investigated using molecular weight enhanced ligands with polyhedral oligomeric silsesquioxanes (POSS) to increase retention in nanofilters.<sup>7</sup> Various hydrogenation reactions also utilizing a rhodium-based homogeneous catalyst in aqueous micellar solutions were investigated for catalyst performance and recovery.<sup>8</sup> Similar procedures have been used to recover the rhodium-based catalyst complex in the hydroformylation of 1-dodecene by using surfactants<sup>9,10</sup> or by using latex polymers as phase transfer agents<sup>11</sup> to capture and separate the catalyst within micelles. However, there still exist complications with catalyst recycling and phase separation

before more serious considerations of surfactant based separation for industrial use can be made. In many of these cases catalyst loss is still too high for an economically justified industrial implementation.

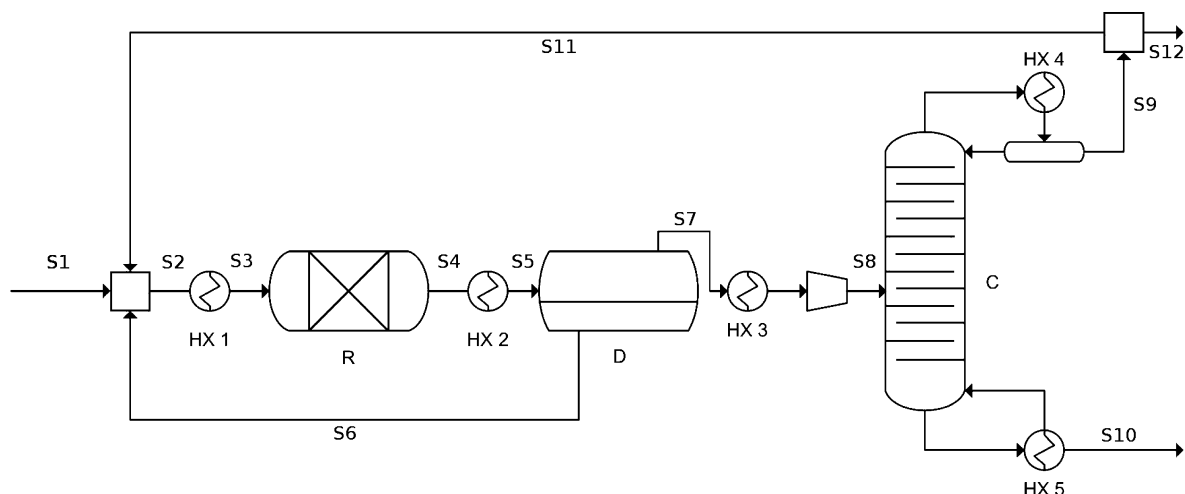
A sizable research effort has focused on the use of temperature sensitive solvent mixtures, called thermomorphic multicomponent solvent systems (TMS), for recovering the catalyst complex from the postreaction mixture.<sup>12</sup> A TMS consists of solvents with varying degrees of polarity that form a homogeneous reaction phase and allow for easy separation in a subsequent cooling stage by then forming two or more phases. The catalyst is expected to be found predominantly in the more polar phase, while the product is expected to be removed in the nonpolar one. These tunable solvent mixtures have been used effectively to recover homogeneous catalysts in various reaction examples.<sup>13–16</sup> A general method for selecting solvents for TMS mixtures was developed using Hansen parameters and the upper critical solution temperature by Behr et al.<sup>15</sup> In this work, the authors first mentioned the use of a TMS consisting of dimethylformamide (DMF) and decane for the hydroformylation of 1-octene and 1-dodecene using rhodium based homogeneous catalysts. Experiments using the reactant 1-dodecene were continued in more detail with DMF/decane solvent mixtures, and promising levels of catalyst recovery<sup>17</sup> and further measures to reduce leaching were successfully achieved.<sup>18</sup> A continuous lab-scale process was designed<sup>19</sup> and implemented<sup>20</sup> using the preferred DMF/decane TMS in equal weight percentages. It is this process, the hydroformylation of

Received: March 3, 2015

Revised: June 10, 2015

Accepted: June 11, 2015

Published: June 25, 2015



**Figure 1.** Flow sheet of the considered, simplified hydroformylation process.

1-dodecene in a TMS of DMF and decane, that is the focus of this work.

Despite these efforts to recover the catalyst, small amounts of catalyst complex and free ligands are lost in the nonpolar product phase; this is otherwise known as catalyst leaching. This small contingent of catalyst is not recovered and is expunged from the process in the final product stream. It is usually imperative to minimize this leaching amount to ensure that an economically feasible process is designed. The solvent composition of the TMS chosen to recover the catalyst and separate out the products plays a major role in the performance of the process. Although these solvent effects on catalyst leaching are known to be important, they have yet to be included in process performance or cost analysis. The major reason for this is the limited reaction kinetic data and catalyst leaching data available. Only one mixture of TMS consisting of 60 wt % decane and 40 wt % DMF is used experimentally in the kinetic data collected by Kiedorf et al.<sup>21</sup> As a result, the optimal reactor trajectories determined by Hentschel et al. are only considered using this fixed composition of the TMS.<sup>22</sup> Additionally, the effect of catalyst leaching is ignored in the optimization; it is assumed that the catalyst and ligands are completely removed in the polar phase in the decanter. In reality this is not the case and different compositions of the TMS can greatly affect reactor performance and catalyst leaching amounts.<sup>18,23,24</sup> These works also reveal that higher molecular weight alkanes and lower temperature phase separation in the decanter have a beneficial effect by reducing catalyst leaching. This data is, regrettably, relatively sparse to use in a process model within the complete allowable design space of solvent selection. Despite this, a simple correlation of the data can be used to estimate the catalyst leaching rates within the process model optimization that is constrained to the conditions of the experiment from which the data was collected. This is useful for the system using a TMS consisting of DMF/decane with a decantation temperature of 25 °C. Also, since reaction kinetics are only available for one TMS composition, a similar correlation for TMS composition dependent conversion is generated. Using these correlations, a process model flow sheet can be designed and optimized with the ratio of DMF to decane left as a decision variable, allowing for solvent-dependent process performance to be investigated.

One issue still remains before the process model can be optimized: a method for determining phase equilibrium. Frequently chemical engineers encounter phase equilibrium calculations when designing new processes and evaluating flow sheets. The general methods of calculating the phase equilibrium are to find a single point based on the conditions of thermodynamic equilibrium. Good initialization values for the start of the iteration are usually required to find a nontrivial solution during optimization, making convergence difficult. To avoid this problem, parameters from an activity coefficient model are fitted to experimental liquid–liquid equilibrium (LLE) data in order to generate a more accurate predictive model. Data generated from this model is then used as input for a linear regression model that will be used to estimate phase partition coefficients in the decanter. This simple equation can then be easily integrated into an optimization problem.

Combining the correlations for conversion and catalyst leaching based on TMS compositions, the linear phase equilibrium model, and a simplified process model allows for the process performance based on the TMS composition to be efficiently evaluated. An optimization problem is then developed based on the total annualized cost of the process over a 3 year period. This model allows one to observe the trade-offs involved between energy requirements, capital costs, reactor performance, and separation performance in the process. When analyzing the results thereof, one can obtain a clearer picture of how catalyst leaching affects the hydroformylation of 1-dodecene leading to the identification of potential areas for improvement.

## ■ PROCESS MODEL

The proposed flow sheet depicted in Figure 1 is a simplified hydroformylation example of the process evaluated in ref 22. A process consisting of a reactor, decanter, and distillation column is used to evaluate the process economics including the catalyst leaching.

The feed ( $S_1$ ) is mixed with two recycle streams and heated to the reaction temperature in a heat exchanger (HX<sub>1</sub>) before being fed into the reactor (R). After the reaction, the mixture is cooled in another heat exchanger (HX<sub>2</sub>) before being fed into the decanter (D) in order to induce phase splitting via the TMS. Two phases are generated here: the denser, catalyst phase, is recycled ( $S_6$ ) and mixed with the feed, while the less

dense, product phase ( $S_7$ ) is fed into a heat exchanger ( $HX_3$ ) to bring the mixture to the bubble point before distillation. This stream ( $S_8$ ) is then fed to the distillation column (C). Here, the product is recovered as the bottoms product ( $S_{10}$ ) and the distillate ( $S_9$ ) containing recovered solvent and reactant is split into a recycle stream ( $S_{11}$ ) and a purge ( $S_{12}$ ). The recycle stream is fed back to the mixer to be combined with the feed and catalyst recycle stream.

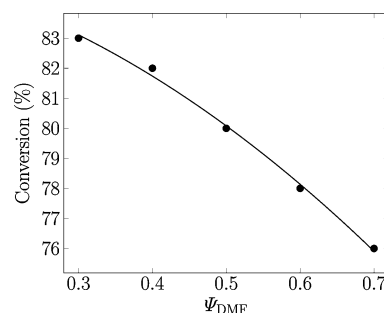
In this model, only one distillation column is required due to our assumption that the reaction does not produce unwanted byproducts. These primary byproducts include 2-methyl-dodecanal, isododecene, and dodecane. The exclusion of these byproducts does not greatly influence the phase separation in the decanter due to their small quantities and/or similarity to other, more abundant molecules in the mixture. This is especially true when using modified UNIFAC Dortmund<sup>25</sup> for LLE calculations as many group contribution methods cannot distinguish between isomers. Two other reactants include carbon monoxide and hydrogen. These gases are an integral part of the reaction and are considered in more detail elsewhere.<sup>22</sup> The separation of unreacted synthesis gas is considered to be relatively simple, and they are assumed to be removed in a flash stage directly after the reactor and are therefore not included in the downstream processing considered in this work.

The effect that the solvent mixture has on the process based on the 3 year averaged total annualized cost (TAC) is then investigated. This consists of the utility cost per kilomole of product and of the capital cost for a process that produces 0.1 kmol of product per hour of operation. In order to find the solvent mixture that leads to the lowest process cost, a nonlinear programming (NLP) optimization problem is developed. In the following sections, more comprehensive descriptions of each unit operation are discussed.

**Reactor.** A simplified black-box reactor governed only by conversion is used instead of the complex reaction network developed by Kiedorf et al.<sup>21</sup> This is due to the lack of kinetic information for differing solvent compositions. The conversion of 1-dodecene is left as a degree of freedom, as the product 1-tridecanal has a pronounced effect on the miscibility of the mixture leaving the reactor and entering the decanter. As seen in ref 24, the LLE of dodecanal/DMF/decane has a much smaller biphasic region than the system *n*-dodecene/DMF/decane. This means that with generation of more product, or higher conversion, a less desirable separation in the decanter occurs which may be associated with higher catalyst loss, higher recycle of product, and/or poor product recovery.

It can also be seen from their work that the conversion of dodecene is dependent on the composition of the TMS, where increasing fractions of DMF lead to slightly lower conversions. In all cases, selectivity remained almost constant. In order to incorporate this data into the model, a quadratic function 1 is used to describe the maximum conversion of dodecene to tridecanal based on the fraction of DMF in the TMS 2 using data collected from Brunsch<sup>23</sup> (Figure 2). All reaction conditions are then fixed to those parameters set by Brunsch: the mass fraction of the TMS is fixed at 0.85, moles of dodecene to rhodium precursor is 1000/1, moles of Biphephos to rhodium precursor is 5/1, and the reaction temperature is constant at 100 °C. The reactor is then otherwise completely described by molar mass balances.

$$X_{\text{dodecene}} = -0.1429\psi_{\text{DMF}}^2 - 0.03714\psi_{\text{DMF}} + 0.8551 \quad (1)$$



**Figure 2.** Correlation of the maximum conversion in the reactor to the mass fraction of DMF in the TMS.

$$\psi_{\text{DMF}} = \frac{\dot{m}_{\text{DMF},4}}{\dot{m}_{\text{DMF},4} + \dot{m}_{\text{decane},4}} \quad (2)$$

$$\dot{n}_{i,3} = \dot{n}_{i,3}, \quad i \in \text{DMF, decane} \quad (3)$$

$$\dot{n}_{\text{tridecanal},4} = (\dot{n}_{\text{tridecanal},3} + \dot{n}_{\text{dodecene},3})X_{\text{dodecene}} \quad (4)$$

$$\dot{n}_{\text{dodecene},4} = (\dot{n}_{\text{tridecanal},3} + \dot{n}_{\text{dodecene},3})(1 - X_{\text{dodecene}}) \quad (5)$$

$$0.85 = w_{\text{DMF},4} + w_{\text{decane},4} \quad (6)$$

$$0.3 \leq \psi_{\text{DMF}} \leq 0.7 \quad (7)$$

**Decanter.** Recovery of the catalyst is perhaps the most economically crucial step and occurs in the decanter. Therefore, the biphasic separation is the most important aspect of this process. The ability to accurately model the decanter during the optimization is highly important.

As in ref 22, the modified UNIFAC Dortmund method is chosen to estimate the activity coefficients of the species in the mixture. For quick predictions of activity and phase equilibrium when experimental data is not available, group combination methods such as modified UNIFAC Dortmund are convenient tools. However, the predictions using the original binary interaction parameters for this system vary considerably for the system DMF/decane/1-dodecene at temperatures higher than around 25 °C and for the system DMF/decane/dodecanal at temperatures in the range -10 to 25 °C. Dodecanal is mentioned here instead of tridecanal due to the lack of experimental data for phase equilibrium behavior including tridecanal. However, data for dodecanal is available from Schäfer et al., who use dodecanal in place of tridecanal for phase equilibrium measurements.<sup>24</sup> Even though dodecanal is investigated instead of tridecanal, the results are not expected to differ significantly when using group methods to predict activity coefficients of the longer aldehyde. If the original binary interaction parameters are used to simulate the model, the phase separation is exaggerated. The miscibility gap for the system DMF/decane/dodecanal is predicted to be larger than it is shown to be experimentally. Using this model, the amount of aldehyde product allowed in the decanter could well exceed the 20% mass constraint. In reality, a maximum of around 15% is maintained to ensure liquid–liquid separation occurs. Therefore, more accurate representation of the phase separation is required.

Since the separation of the TMS in the decanter is of paramount importance for the overall process performance, the inaccuracies in the liquid–liquid equilibrium prediction need to be overcome. To do this, new binary interaction parameters are fitted to experimental data from ref 24. Some of these

Table 1. Regressed (Italics) Binary Interaction Parameters for Use with Modified UNIFAC Dortmund

	$A_{ij}$			
	CH <sub>2</sub>	HCO	DMF	CH <sub>2</sub> =CH
CH <sub>2</sub>	0	639.581 004	871.437 927	189.66
HCO	-504.634 64	0	-549.813 02	202.49
DMF	114.342 456	92.889 318	0	-55.044 021
CH <sub>2</sub> =CH	-95.418	476.25	1033.737 82	0

parameters are originally fitted by Ye,<sup>26</sup> but the binary interaction parameters for HCO–DMF and HCO–CH<sub>2</sub> are modified slightly to more accurately describe the phase behavior of the DMF/decane/dodecanal system. Phase diagrams comparing experimental data to the predictions made using the new parameters (Tables 1 and 2) are shown in Figures 3 and 4. It can be clearly seen that the newly fit parameters provide good predictability of the phase splitting behaviors.

Table 2. Regressed (Italics) Binary Interaction Parameters for Use with Modified UNIFAC Dortmund

	$B_{ij}$			
	CH <sub>2</sub>	HCO	DMF	CH <sub>2</sub> =CH
CH <sub>2</sub>	0	0	-0.951 592 9	-0.272 32
HCO	0	0	0	0
DMF	-0.754 095 2	0	0	-0.357 397 4
CH <sub>2</sub> =CH	0.061 708	0	-2.159 510 5	0

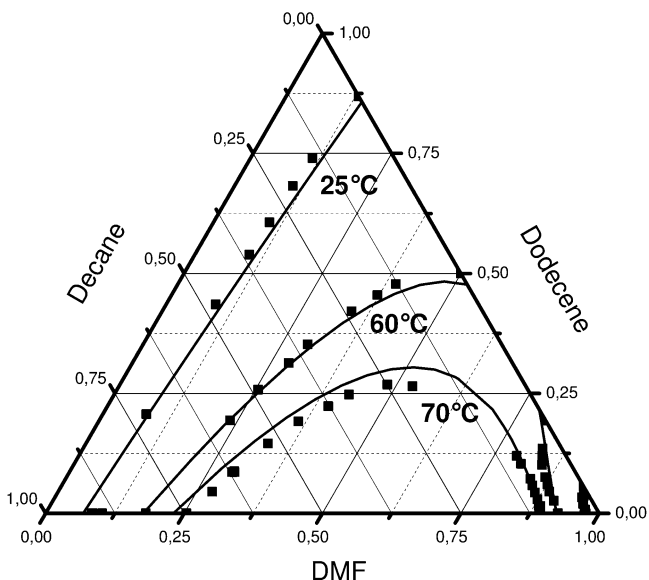


Figure 3. Ternary LLE diagram for the system DMF/decane/1-dodecene with regressed binary interaction parameters. The solid lines represent the predicted LLE, and the points represent the experimental data collected by Schäfer et al.<sup>24</sup>

An issue encountered by Hentschel et al.<sup>22</sup> is the difficulty in calculating the phase separation during the optimization. Their solution for this problem is to calculate partition coefficients  $\theta$  using regressed second order equations with conversion as the independent variable. This correlation is found to be very agreeable with the phase separation as calculated using modified UNIFAC Dortmund (however, without fitted parameters). A similar idea is used in this work, but instead of basing the model on conversion, it is based on the

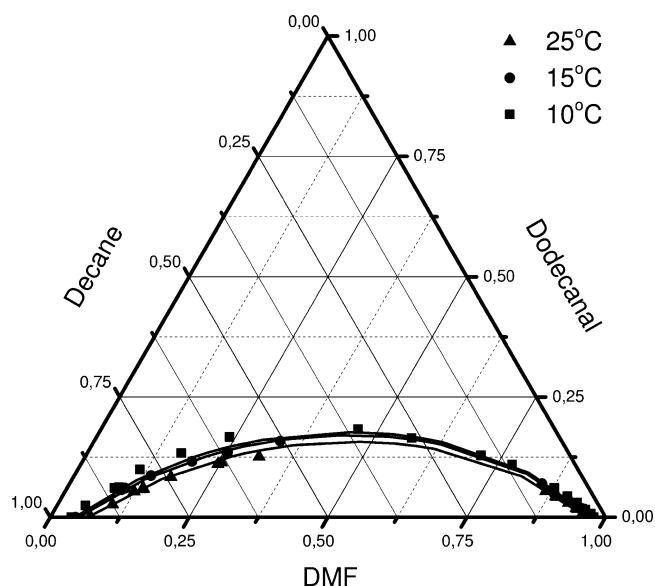


Figure 4. Ternary LLE diagram for the system DMF/decane/dodecanal with regressed binary interaction parameters. The solid lines represent the predicted LLE, and the points represent the experimental data collected by Schäfer et al.<sup>24</sup>

composition of the inlet to the decanter, allowing for greater flexibility in the TMS composition. This new model is also linear with respect to its coefficients. The benefit of using a linear regression model lies in its simplicity and uncomplicated differentiation. Such models can then be smoothly incorporated into an optimization problem, thus averting the difficulties inherent in phase equilibrium convergence.

Since there are only four components considered in this study, four different linear regression models (eq 9) are needed: one for each partition coefficient. By using partition coefficients, the number of moles of each species in each phase as well as the total size of each phase can be calculated. These coefficients are to be dependent on the composition of the mixture entering the decanter. The obvious choices for the independent variables are the mole fractions of this mixture.

$$\theta_i^\alpha = \frac{n_i^\alpha}{n_i^\alpha + n_i^\beta}, \quad i \in \text{COM} \quad (8)$$

$$\theta_i^\alpha = f_i(\hat{x}_3), \quad i \in \text{COM} \quad (9)$$

To generate data for the linear regression model, the phase equilibrium for 10 000 randomly initialized compositions are calculated at 25 °C using the rate based method developed by Steyer et al.<sup>27</sup> Of these, roughly 51% are found to be within the biphasic region. Using these biphasic points, the four linear regression models are calculated using the statistics toolbox in MATLAB (version R2012b). This is a very convenient method for removing insignificant terms or adding significant terms to

the regression model. It turns out that the linear regression model predicts the partition coefficients very well for all mixtures that do not contain coefficients with values close to zero. It must be noted that only compositions within the biphasic region should be used with this model. Otherwise, obscure values larger than 1 or less than 0 for the partition coefficients will be returned. As long as the system being investigated is well-defined and constrained to the biphasic region, no issues should be encountered. Using these models, rather accurate results are generated that provide all of the required information to describe the liquid–liquid equilibrium. The linear regression models and the parameters used in this study are found in the Supporting Information.

As seen in Table 3, when partition coefficients close to zero are removed from the regression, the average percent error

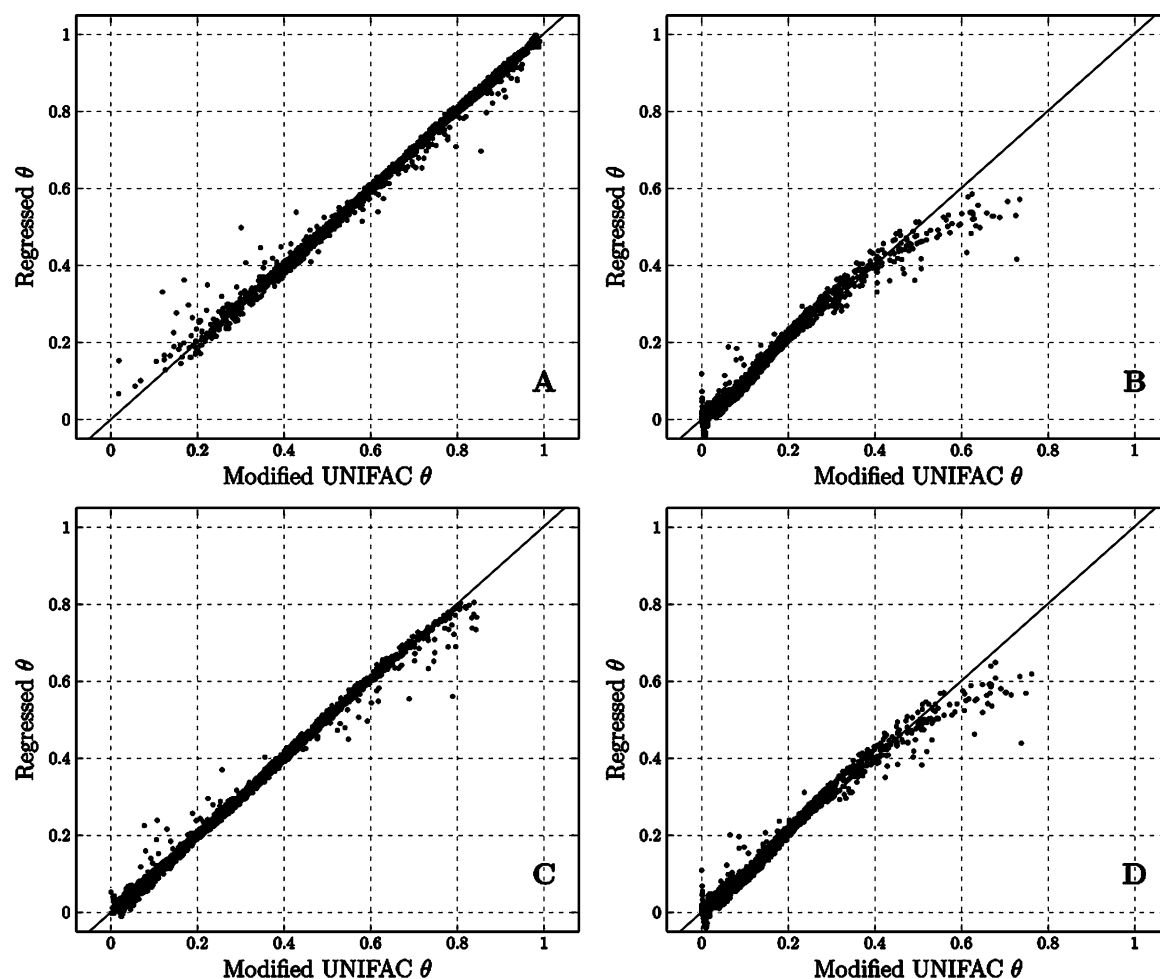
**Table 3. Linear Regression Model Errors for Acceptable Ranges of  $\theta$  (eq 9)**

	$\theta$ range			
	0.0–1.0	0.01–1.0	0.02–1.0	0.03–1.0
MAPE (%)	22.13	7.92	4.92	3.48
MAE	0.0068	0.0056	0.0049	0.0046
data points	5056	4586	3965	3414

decreases substantially while the relative error changes less drastically. This simply means that, for components with very low partition coefficients, the error can be quite large, which is more so reflected in the percent error than in the absolute error. Figure 5 compares the partition coefficients as predicted using the linear regression model with the data calculated using modified UNIFAC Dortmund with fitted parameters. In Figure 5, the complete set of data points within the biphasic region is represented. It can be seen that a reasonable number of outliers is present, but overall the linear model accurately reflects the phase splitting behavior of the system. It is interesting to note that these outliers represent high fractions of nonpolar components in the polar phase.

One caveat with this method of determining the phase equilibrium is that it can only be calculated for the temperature with which the data was generated. In future models, it may be possible to incorporate temperature into the regression model. This would be worthwhile if enough data existed for catalyst leaching at different temperatures.

Since two phases are generated in the decanter, the recycle and product phases need to be accurately defined. This is done using the density of each phase. These densities are calculated using the summation of each component density multiplied by its respective weight fraction. The lower density phase is assigned as the product phase. The more dense phase is the more polar phase containing the majority of polar solvent and



**Figure 5.** Parity plot of partition values for the quaternary system consisting of DMF (A), decane (B), tridecanal (C), and dodecene (D) at 25 °C.

is recycled to the reactor. This procedure is necessary because the linear regression model does not return partition coefficients specific to one of the phases. The decanter is then described by the following system of equations in addition to the linear regression models mentioned above:

$$\dot{n}_{i,\text{dec}_1} = \theta_i^\alpha \dot{n}_{i,5}, \quad i \in \text{COM} \quad (10)$$

$$\dot{n}_{i,\text{dec}_2} = (1 - \theta_i^\alpha) \dot{n}_{i,5}, \quad i \in \text{COM} \quad (11)$$

$$\rho_i = a_{p,0,i} + a_{p,1,i} T, \quad i \in \text{COM} \quad (12)$$

$$\rho_{\text{dec}_j} = \sum_i w_{i,\text{dec}_j} \rho_i, \quad i \in \text{COM}, j \in \text{PHASE} \quad (13)$$

$$\rho_{\text{dec}_1} \geq \rho_{\text{dec}_2} \quad (14)$$

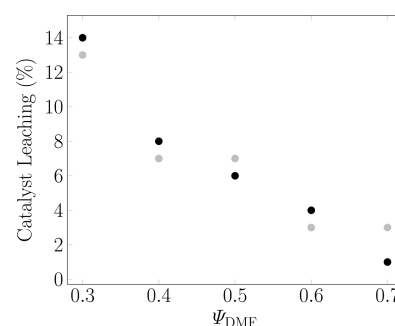
$$\dot{n}_{i,6} = \dot{n}_{i,\text{dec}_1}, \quad i \in \text{COM} \quad (15)$$

$$\dot{n}_{i,7} = \dot{n}_{i,\text{dec}_2} = \dot{n}_{i,5} - \dot{n}_{i,6}, \quad i \in \text{COM} \quad (16)$$

**Catalyst Leaching.** The purpose behind using a TMS is to economically recover the catalyst by inducing phase separation by cooling after the reaction is complete. A small amount of catalyst complex and excess ligands in solution that remain in the less-dense product phase are deactivated in the distillation column and are removed from the process as part of the product stream. Since this amount is quite small, in the range of a few parts per million or less, its effect on the purity of the tridecanal product is negligible. In order to evaluate the effectiveness of the chosen TMS system, the effect of catalyst leaching must be included in the optimization problem. This is especially true due to the high prices of these ligands.

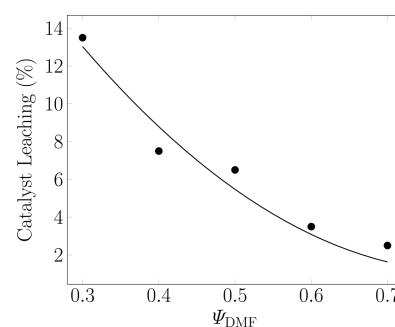
A kilogram price of \$38.5K is adopted for the Biphephos ligand based on past purchase information from 2010 (MOLISA GmbH). The cost of the rhodium(I) dicarbonyl acetylacetonate precursor is estimated using the spot price of rhodium at the time of the Biphephos acquisition, \$2,648 per troy ounce (spring 2010). Of course, the cost of the finished rhodium precursor will have a higher cost than the spot metal price, but this is not considered here. Care must also be taken when using rhodium due to its highly volatile price as this may have a significant effect on process economics.

As previously mentioned, Hentschel et al.<sup>22</sup> do not consider catalyst leaching in their reactor and process optimization problem. It is assumed that the catalyst is completely recovered in the polar phase and recycled to the reactor. It is known, however, that catalyst leaching occurs for this system.<sup>23,24</sup> The amount of catalyst loss also changes with the composition of the TMS, and perhaps on the amount of product produced. Although there is some information about the catalyst leaching available, it is lacking in its depth for greater analysis of this problem. Figure 6 depicts relative levels of leaching for rhodium and ligands for various TMS compositions for the reaction system defined in ref 23. With increasing levels of polar solvent in the TMS, the amounts of leaching for both rhodium and ligands decrease at around the same rate. Assuming that leaching of the rhodium can be combined with that of the ligands, a correlation can be developed with equal rates of leaching for both catalyst components. The average of their respective leaching amounts is then considered to be the overall magnitude of catalyst leaching. This average was then correlated to the mass fraction of DMF in the TMS and a very good fit was observed (see Figure 7). With more data, it



**Figure 6.** Catalyst leaching represented by rhodium (black) and phosphorus (gray) loss relative to TMS composition.

may be necessary to fit separate correlations for rhodium and ligand leaching levels.



**Figure 7.** Correlation between combined catalyst leaching and the mass fraction of DMF in the TMS.

The correlation is represented by a quadratic equation (17) and represents catalyst leaching based on the mass fraction of DMF in the TMS. A linear representation is unsuitable in that it greatly underestimates catalyst loss for higher levels of DMF in the solvent composition. This correlation is used in conjunction with the total mass of the catalyst used in the process based on the number of moles of dodecene entering the reactor. Using a new variable, the percent of catalyst leaching, a well-defined value for the catalyst loss can be incorporated into the cost optimization of the process.

$$w_{\text{cat,leaching}} = 0.4643\psi_{\text{DMF}}^2 - 0.7493\psi_{\text{DMF}} + 0.3133 \quad (17)$$

**Distillation Column.** The nonpolar phase leaving the decanter consisting of all four components is fed to the distillation column. The product is to be separated at a high purity (at least 99.9%) from the solvents and will be captured as the bottoms,  $S_{10}$ , due to its low vapor pressure. The other solvents are condensed and form the distillate. A small purge is separated from the distillate and the rest,  $S_{11}$ , is recycled to the mixer and combined with the feed stream,  $S_1$ , and the catalyst recycle stream leaving the decanter,  $S_6$ .

The energy requirements for the distillation column are governed by the reboiler and condenser duties related to the reflux and boilup ratios. The column flows are modeled using the Underwood correlation, assuming constant molar overflow and constant relative volatilities. The temperature of the column is restricted to a maximum of 180 °C because the tridecanal product begins to degrade at such temperatures. Since the normal boiling point of tridecanal lies well above 180 °C, the column is operated under vacuum and is assumed have a pressure drop of 50%. This column differs from Hentschel's<sup>22</sup>

solvent column in that the pressure is lowered to more accurately reflect the desired temperature in the bottoms. The bottoms pressure should not be higher than 90 mbar, leaving the pressure at the top of the column to be 60 mbar. Vapor pressure correlations (eq 18) are used to calculate relative volatilities.<sup>28</sup> The vapor pressure parameters used in these correlations are provided in the Supporting Information.

$$p_{\text{vap},i} = 10^{(a_0 + a_1/T + a_2 \log(T) + a_3 T + a_4 T^2)(133.322 \times 10^{-6})},$$

$$i \in \text{COM} \quad (18)$$

Due to our simplifications, the relative volatilities can be calculated using a simple vapor pressure ratio, where component  $j$  is the heavy key.

$$\alpha_{i,j}(T) = \frac{p_{\text{vap},i}(T)}{p_{\text{vap},j}(T)} \quad (19)$$

The mean relative volatility is calculated as the geometric mean of the relative volatility of the distillate (D) and that of the bottoms (B). The feed is considered to be a saturated liquid ( $q = 1$ ), and therefore the distillation feed stream,  $S_8$ , requires heating after leaving the decanter,  $S_7$ .

$$\bar{\alpha}_{i,j} = \sqrt{\alpha_{i,j}(T_D) \alpha_{i,j}(T_B)} \quad (20)$$

Using key component outlet stream data, the minimum number of trays for the column can be calculated with the Fenske equation.

$$N_{\text{min}} = \frac{\log\left(\frac{\dot{n}_{D,LK} \dot{n}_{B,HK}}{\dot{n}_{B,LK} \dot{n}_{D,HK}}\right)}{\log(\bar{\alpha}_{LK,HK})} \quad (21)$$

Now, combining the molar mass balance of the column with the minimum number of trays, the molar flow rates for all components in the outlet streams of the column can be derived.

$$\dot{n}_{B,i} = \frac{\dot{n}_{in,i}}{1 + \frac{\dot{n}_{D,HK} \bar{\alpha}_{i,HK} - N_{\text{min}}}{\dot{n}_{B,HK}}} \quad (22)$$

$$\dot{n}_{in,i} = \dot{n}_{D,i} + \dot{n}_{B,i} \quad (23)$$

In this model, the amount of key component recovered in each outlet stream is left as an optimization variable. Recovery of tridecanal must be at least 80% and no greater than 99.999% in the bottoms. The maximum recovery is expected to be high due to the large difference in volatilities of tridecanal and dodecene. The amounts of key component in the distillate and bottoms are governed by the recovery ratios  $\zeta_{\text{dodecene}}$  and  $\zeta_{\text{tridecanal}}$  for the light and heavy keys, respectively. As in ref 22, the sum of the recovery ratios is 1.

$$\dot{n}_{D,LK} = \zeta_{\text{dodecene}} \dot{n}_{in,\text{dodecene}} \quad (24)$$

$$\dot{n}_{D,HK} = \zeta_{\text{tridecanal}} \dot{n}_{in,\text{tridecanal}} \quad (25)$$

$$1 = \zeta_{\text{tridecanal}} + \zeta_{\text{dodecene}} \quad (26)$$

$$0.8 \leq \zeta_{\text{dodecene}} \leq 0.99999 \quad (27)$$

**Optimization.** The goal of this work is to identify the cost optimal process considering catalyst leaching for the previously described model. The major trade-offs are the lowered conversion in the reactor and larger recycle streams competing

with lowered catalyst loss when using higher levels of DMF in the TMS. Using cost functions is the best way to ensure that the best compromise is located. Cost functions are taken from Guthrie,<sup>29</sup> and sizing estimations are made using methods developed by Biegler.<sup>30</sup> The objective function is based on the 3 year total annualized cost of production. This includes the depreciation of the capital equipment, all utilities, feedstock, and makeup solvent and catalyst. Two slightly different optimizations will take place: one with catalyst lost based on rhodium alone and a second with a combined catalyst loss considering both rhodium and ligand leaching. This is done to show the importance of ligand loss compared to only focusing on rhodium, as previous works have done. Optimization variables in this process include the feed composition, the composition of the TMS in the reactor, recovery fractions of the distillate and bottoms, and the reflux ratio of the distillation column.

All optimization problems are implemented in AMPL and optimized using the solver CONOPT 3.14 V on a PC with one Intel Core i5-3570 CPU at 3.40 GHz, a memory of 8 GB, and running on the Kubuntu 12.04 operating system.

$$\text{Obj} = \min_{U(t)} \left( \frac{C_{\text{capital}}/3 + C_{\text{utilities}} + C_{\text{feedstock}}}{\dot{n}_{10,\text{tridecanal}}} \right),$$

$$U = \dot{n}_1, \xi_{\text{dodecene}}, \Psi_{\text{DMF}}, R_{\text{factor}} \quad (28)$$

s.t. the following: reactor, eqs 1–5; solvent composition, eqs 6 and 7; decanter, eqs 8–16; catalyst leaching, eq 17; distillation column, eqs 18–27; flow sheet constraints, eqs S2–S24 in the Supporting Information; utilities, eqs S25–S37 in the Supporting Information; sizing, eqs S38–S68 in the Supporting Information; cost models, eqs S69–S76 in the Supporting Information; product specifications,  $\dot{n}_{\text{tridecanal},10} = 0.1$  kmol/min.

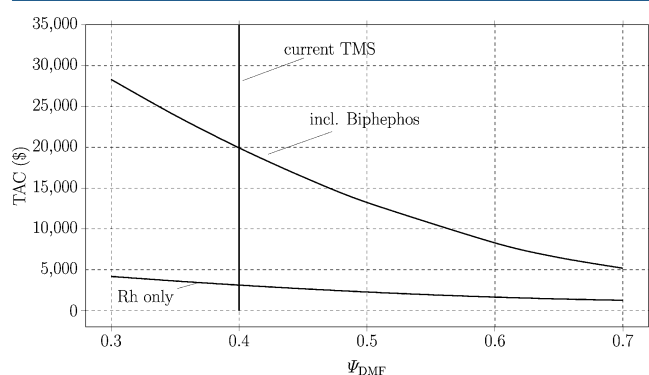
## DISCUSSION AND RESULTS

In both optimization problems, the optimal operating point is found to be the point with the highest allowable DMF mass fraction of 0.7. The exceptionally high cost due to catalyst leaching is responsible for the bulk of the process cost. In order to reduce the amount of catalyst lost, larger amounts of polar DMF are used in the TMS to reduce leaching. The lowered conversion and higher recycle costs occurring with higher DMF to decane ratios lead to higher capital and utility costs that are minuscule in comparison to the saved costs due to increased catalyst retention. This proves that catalyst loss and its relevance to solvent selection are the most important criteria for further study.

**Catalyst Leaching.** Optimizing this process based on the composition of the TMS, the feed or makeup streams, the recovery fraction, and reflux ratio, the optimal point is found to be the process with the maximum allowable DMF with 70% by mass in the TMS mixture. This is not surprising considering the high cost of the catalyst in comparison to the other utility costs; with  $\Psi_{\text{DMF}}$  equal to 0.3, the catalyst is 97.6% of the utility cost and with  $\Psi_{\text{DMF}}$  equal to 0.7, its fraction of the utility cost reduces only to 86.8%. This is met, however with a substantial decrease in the overall process cost. By increasing  $\Psi_{\text{DMF}}$ , the total cost of the process decreases significantly.

It is useful at this point to compare the performance of the solvent mixture examined by Hentschel<sup>22</sup> and Kiedorf<sup>21</sup> where  $\Psi_{\text{DMF}}$  is equal to 0.4. The optimal process cost based on rhodium loss is then equal to \$3,122/kmol product. As  $\Psi_{\text{DMF}}$  is

increased to 0.7, the total cost drops significantly to \$1,260/kmol product, a reduction of 59.6%. For more comparisons, the optimal process costs for TMS composition ratios ranging from 0.3 to 0.7 are shown in Figure 8. The results of the second



**Figure 8.** Total cost per kilomole of tridecanal produced with respect to the TMS composition,  $\Psi_{\text{DMF}}$ .

optimization including ligand leaching show a much more pronounced effect on the process cost. One sees that the cost increases by at least an order of magnitude when ligands are included in the cost function (Figure 8). It is again possible to make a direct comparison as before. The cost of this process considering ligand loss is \$19,223/kmol product with  $\Psi_{\text{DMF}}$  equal to 0.4, and the cost reduces to \$5,191/kmol product with  $\Psi_{\text{DMF}}$  increasing to 0.7. This means that a 73.9% reduction in total process cost could be expected simply by using a different composition of the TMS.

Much discussion takes place concerning the loss of metal catalyst due to its cost, but it seems that the same if not more attention should be given to the ligand. Not only does the ligand follow similar leaching patterns as rhodium, it has a higher concentration in the mixture (5 times the mole number of metal) and about 3 times the molecular mass. This means that for every unit mass of rhodium lost due to leaching roughly 15 times as much ligand is lost. As seen in Table 4, the cost of

**Table 4. Cost Breakdown of the Optimal Points ( $\Psi_{\text{DMF}} = 0.7$ ) for Catalyst Costs Excluding and Including Biphephos (BPP) Leaching Cost<sup>a</sup>**

	rhodium	dodecene	utilities/ solvents	biphephos	TAC
excl BPP	45.4%	53.0%	1.6%	—	\$1,260/kmol
incl BPP	11.0%	12.8%	0.4%	75.8%	\$5,191/kmol

<sup>a</sup>Values given are percent TAC.

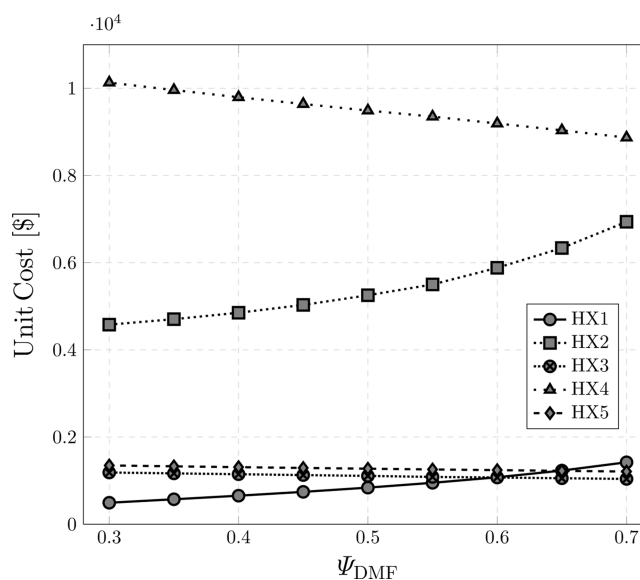
Biphephos at the optimal point is about 76% of the total process TAC while rhodium loss accounts for only 11%. If the cost of Biphephos is not considered, then rhodium loss is quite significant but only 45% of the total cost. It would seem more focus should be made on ligand retention in addition to rhodium recovery.

It is not unexpected that more polar solvent would lead to better separation of the catalyst; this was previously discussed by Brunsch.<sup>23</sup> The degree to which a simple change in the TMS would affect the process cost is unanticipated. It shows that, before considering detailed kinetics and downstream processing, the economic bottleneck or linchpin of the process should

be thoroughly investigated. These results show that maximizing conversion, even for the simplified reaction here, would only benefit the process insignificantly if the catalyst loss problem is not addressed beforehand.

Optimizations executed excluding catalyst costs only varied in total annualized cost of around 0.51%. This accounts for the increased conversion when higher levels of decane are used in the TMS. In this scenario, the cost optimal process uses a  $\Psi_{\text{DMF}}$  equal to 0.3 and a TAC of \$688 and the cost continues to rise with increasing levels of DMF to a maximum of \$691 with  $\Psi_{\text{DMF}}$  equal to 0.7. The majority of this cost is the dodecene feed. As seen above, the overwhelming impact of catalyst leaching on the cost of the process makes the increased capital and utility costs associated with lower conversion negligible. If we optimize the process without taking the catalyst leaching into account, the optimal configuration provided may be quite the opposite.

**Capital Costs.** Although accounting for a very small percentage of the overall process costs (0.0083% at  $\Psi_{\text{DMF}} = 0.3$  to 0.054% at  $\Psi_{\text{DMF}} = 0.7$ ), it is interesting to see how the individual capital costs of each unit vary with TMS composition. In Figure 9 the unit capital costs for the five

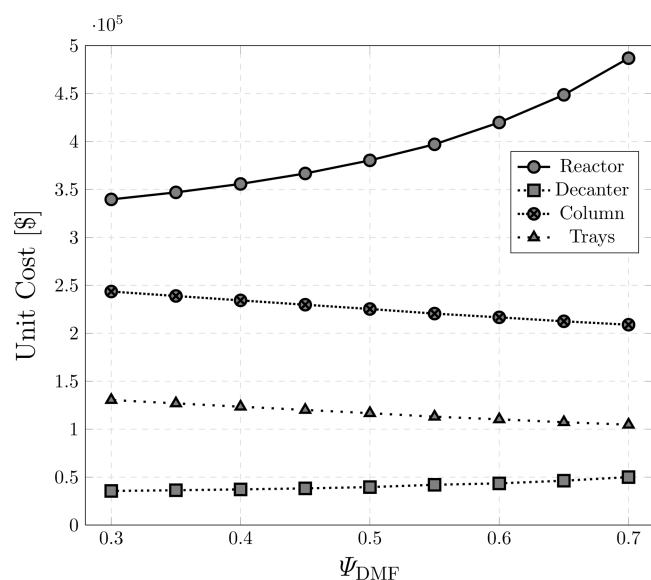


**Figure 9.** Total capital cost for the five heat exchangers for various TMS compositions.

heat exchangers are shown, and in Figure 10 the unit capital costs for the reactor, decanter, distillation column, and trays are shown. Increasing amounts of polar solvent lead to a larger catalyst recycle stream ( $S_6$ ) which is seen in the increasing size of the reactor and the heat exchangers before and after the reactor. The size of the decanter increases only slightly. With a higher fraction of the total mixture in the decanter being recycled, the burden on the distillation column decreases slightly. This is somewhat expected as the required production amount of tridecanal leads to a fairly constant bottoms stream. As the separation in the decanter improves, less DMF enters the distillation column, which decreases the overall flow rate and hence its size. This is also seen in the decrease in size of the condenser, the largest heat exchanger in the flow sheet.

Considering the extremely low capital cost of this process, any increase in process size or extra units used that helps to improve catalyst retention or decrease the amount lost would

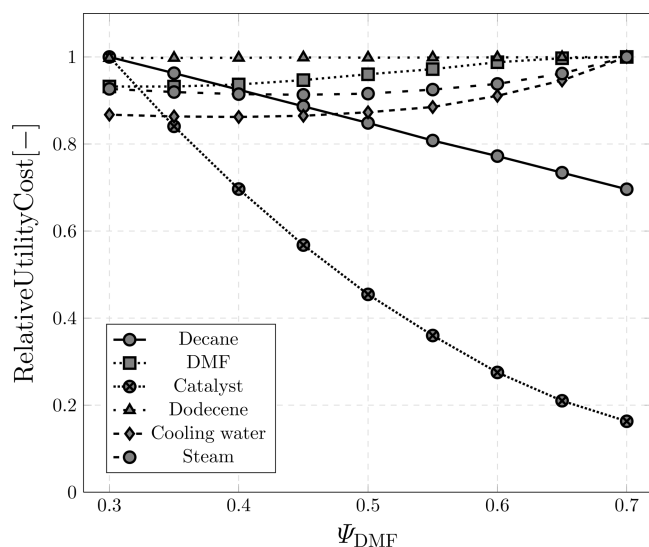




**Figure 10.** Total capital cost for the reactor, decanter, distillation column shell, and distillation trays for various TMS compositions.

be beneficial. This may mean longer residence times in the reactor with lower amounts of catalyst used per amount product, thereby increasing the total volume of each unit and the utilities used. In contrast to the reduced catalyst leaching, these increased costs would be very small in comparison. It is therefore recommended to investigate such schemes in future process designs utilizing this catalyst and TMS.

**Utility Costs.** In Figure 11 the relative utility costs for the feedstock, cooling water, steam, and catalyst can be seen. The



**Figure 11.** Relative utility costs for various TMS compositions.

largest variation can be seen in the relative catalyst cost. As this is by far the largest cost, this large deviation has a sizable impact on the overall cost of the utilities. As  $\Psi_{DMF}$  increases, the cost of the catalyst significantly decreases in a similar manner as the total annualized process cost. All other utilities increase with increasing values of  $\Psi_{DMF}$  except for decane, which understandably decreases. Due to the large increase in the size of the reactor, there is also a parallel increase in heating and cooling duties thereby increasing the amount of steam and cooling

water required. Only the amount of dodecene remains relatively constant as DMF is increased. This is due to the required product constraint and that most of the unconverted dodecene is recycled back into the reactor. However, when considering byproduct formation, most of the unconverted dodecene would be isomerized in the reactor forming isododecene. As shown by Hentschel,<sup>22</sup> recycling isododecene has the benefit of reducing the amount of fresh dodecene that is isomerized increasing the overall selectivity and conversion to tridecanal. This means that, in a real process at steady state, the required amount of dodecene in the feed would not vary much with changing TMS compositions.

The recovery fraction of dodecene in the distillate is optimal at very high ratios. With  $\Psi_{DMF}$  equal to 0.3, the recovery fraction of dodecene is 0.998 801 and rises to 0.999 463 with  $\Psi_{DMF}$  equal to 0.7. This basically means that very little tridecanal is found in the distillate. This shows that using the recovery rate as an optimization variable is insignificant. It would be a good assumption to model the column as an indirect separation with a nondistributing bottoms product, simplifying the model further. The effect of the reflux ratio is to decrease the size of the column by finding the balance between the diameter and the height of the column. In this case, the reflux ratio steadily increases as  $\Psi_{DMF}$  increases, moving from 0.065 to 0.074. The reflux rate is thus more important in this process than the recovery rate. This is, however, only slightly significant as both are rather insensitive in the overall process optimization.

## CONCLUSION

A complex process is simplified to focus on the importance that catalyst recovery has on the hydroformylation of 1-dodecene. By incorporating catalyst leaching into the process design problem using correlations from limited data, significant effects of catalyst loss on the total annualized costs are able to be seen. A similar method is implemented for the reaction system by replacing the complex reaction kinetics with a correlation taken from the same experimental data. The complex phase equilibrium behavior is also successfully represented using linear regression models to simplify the phase separation calculation during optimization. These allow for a much faster and simpler optimization problem focusing on the solvent composition. Results from the optimization show a strong dependency of process economics on the TMS composition and suggest that cost reduction strategies should focus on reducing the cost of the catalyst or decreasing the incidence of leaching to less than negligible levels.

The solvent composition has a large role in process costs. In fact, almost all other areas of process optimization such as reactor design bring about negligible changes in regard to process cost at the current levels of catalyst leaching. This is clearly seen when referring to the optimization problem excluding catalyst leaching costs on the process where an insignificant change in process costs was seen. Increasing the solvent to the maximum allowable amount of DMF that would still maintain a biphasic mixture upon cooling would be the most effective and immediate measure at reducing the cost of the process. Other recommendations include increasing the size of the reactor, and in practice all other process elements, in order to increase the residence time of the mixture during the reaction. This would allow for a reduction in the catalyst concentration and overall amount used which would lead to a reduction in leaching. Due to the extreme difference between

catalyst and capital costs, it would be beneficial for the economic viability of the process to decrease operating costs at the expense of increasing investment costs.

From experimental data, it is known that other changes can be made to better recover the catalyst. The effect of lowering the temperature in the decanter has a larger positive influence on catalyst leaching than on the degree of phase separation. By lowering the temperature used for phase separation, significant levels of catalyst may become recoverable. However, more data is required before a reasonable model or even correlation can be made. Using different temperatures would also require a better model for phase separation, such as kriging or using artificial neural networks; the linear regression model included is only suitable for a single temperature. Using different compositions that are not constrained to the data used in this work would also lead to better understanding about the catalyst leaching problem. It may be that higher compositions of overall solvent (more than the 85% total mass in the reaction mixture) may reduce leaching further. It would also be worthwhile to investigate improved methods of choosing optimal component solvents for the TMS in order to find mixtures that better facilitate recovery of the catalyst. This would require detailed studies into the solubility of the catalyst complex or ligands in different mixtures. As it clearly seems to be the most important aspect of this process, this course of action is highly recommended.

## ■ ASSOCIATED CONTENT

### 📄 Supporting Information

Nomenclature, pure substance properties, process flow sheet model constraints, utility and unit size models, cost functions including capital investment, utility, and materials. The Supporting Information is available free of charge on the ACS Publications website at DOI: 10.1021/acs.iecr.5b00795.

## ■ AUTHOR INFORMATION

### Corresponding Author

\*E-mail: sundmacher@mpi-magdeburg.mpg.de.

### Notes

The authors declare no competing financial interest.

## ■ ACKNOWLEDGMENTS

This work was conducted in part in cooperation with the Collaborative Research Centre "Integrated Chemical Processes in Liquid Multiphase Systems". The financial support from the Deutsche Forschungsgemeinschaft (DFG) is sincerely acknowledged (TRR 63).

## ■ REFERENCES

- (1) Bohnen, H.-W.; Cornils, B. Hydroformylation of alkenes: An industrial view of the status and importance. *Adv. Catal.* **2002**, *47*, 1–64.
- (2) Cole-Hamilton, D. J. Homogeneous catalysis—new approaches to catalyst separation, recovery, and recycling. *Science* **2003**, *299*, 1702–1706.
- (3) Hintermair, U.; Zhao, G.; Santini, C. C.; Muldoon, M. J.; Cole-Hamilton, D. J. Supported ionic liquid phase catalysis with supercritical flow. *Chem. Commun.* **2007**, 1462–1464.
- (4) Song, K.-C.; Baek, J. Y.; Bae, J. A.; Yim, J.-H.; Ko, Y. S.; Kim, D. H.; Park, Y.-K.; Jeon, J.-K. Octene hydroformylation by using rhodium complexes tethered onto selectively functionalized mesoporous silica and in situ high pressure IR study. *Catal. Today* **2011**, *164*, 561–565.
- (5) Fang, J.; Jana, R.; Tunge, J. A.; Subramaniam, B. Continuous homogeneous hydroformylation with bulky rhodium catalyst com-

plexes retained by nano-filtration membranes. *Appl. Catal., A* **2011**, *393*, 294–301.

- (6) Xie, Z.; Fang, J.; Subramaniam, B.; Maiti, S. K.; Snavely, W.; Tunge, J. A. Enhanced hydroformylation by carbon dioxide-expanded media with soluble Rh complexes in nanofiltration membrane reactors. *AIChE J.* **2013**, *59*, 4287–4296.

- (7) Janssen, M.; Wilting, J.; Müller, C.; Vogt, D. Continuous rhodium-catalyzed hydroformylation of 1-octene with polyhedral oligomeric silsesquioxanes (POSS) enlarged triphenylphosphine. *Angew. Chem., Int. Ed.* **2010**, *49*, 7738–7741.

- (8) Schwarze, M.; Milano-Brusco, J.; Stempel, V.; Hamerla, T.; Wille, S.; Fischer, C.; Baumann, W.; Arlt, W.; Schomäcker, R. Rhodium catalyzed hydrogenation reactions in aqueous micellar systems as green solvents. *RSC Adv.* **2011**, *1*, 474–483.

- (9) Schwarze, M.; Schmidt, M.; Nguyen, L.; Drews, A.; Kraume, M.; Schomäcker, R. Micellar enhanced ultrafiltration of a rhodium catalyst. *J. Membr. Sci.* **2012**, *421–422*, 165–171.

- (10) Hamerla, T.; Rost, A.; Kasaka, Y.; Schomäcker, R. Hydroformylation of 1-dodecene with Water-Soluble Rhodium Catalysts with Bidentate Ligands in Multiphase Systems. *ChemCatChem* **2013**, 1854–1862.

- (11) Nowothnick, H.; Rost, A.; Hamerla, T.; Schomäcker, R.; Müller, C.; Vogt, D. Comparison of phase transfer agents in the aqueous biphasic hydroformylation of higher alkenes. *Catal. Sci. Technol.* **2013**, *3*, 600–605.

- (12) Behr, A.; Fängewisch, C. Temperature-Dependent Multi-component Solvent Systems—An Alternative Concept for Recycling Homogeneous Catalysts. *Chem. Eng. Technol.* **2002**, *25*, 143–147.

- (13) Behr, A.; Roll, R. Temperaturgesteuerte Mehrkomponenten-Lösungsmittelsysteme für homogene Übergangsmetallkatalysierte Reaktionen. *Chem. Ing. Technol.* **2005**, *77*, 748–752.

- (14) Behr, A.; Roll, R. Hydroaminomethylation in thermomorphic solvent systems. *J. Mol. Catal. A: Chem.* **2005**, *239*, 180–184.

- (15) Behr, A.; Henze, G.; Johnen, L.; Awungacha, C. Advances in thermomorphic liquid/liquid recycling of homogeneous transition metal catalysts. *J. Mol. Catal. A: Chem.* **2008**, *285*, 20–28.

- (16) Behr, A.; Johnen, L.; Neubert, P. A sustainable route from the renewable myrcene to methyl ethers via direct hydroalkoxylation. *Catal. Sci. Technol.* **2012**, *2*, 88–92.

- (17) Behr, A.; Brunsch, Y.; Lux, A. Rhodium nanoparticles as catalysts in the hydroformylation of 1-dodecene and their recycling in thermomorphic solvent systems. *Tetrahedron Lett.* **2012**, *53*, 2680–2683.

- (18) Brunsch, Y.; Behr, A. Temperature-controlled catalyst recycling in homogeneous transition-metal catalysis: minimization of catalyst leaching. *Angew. Chem., Int. Ed.* **2013**, *52*, 1586–1589.

- (19) Behr, A.; Witte, H.; Zagajewski, M. Scale-up durch Miniplant-Technik: Anwendungsbeispiele aus der homogenen Katalyse. *Chem. Ing. Technol.* **2012**, *84*, 694–703.

- (20) Zagajewski, M.; Behr, A.; Sasse, P.; Wittmann, J. Continuously operated miniplant for the rhodium catalyzed hydroformylation of 1-dodecene in a thermomorphic multicomponent solvent system (TMS). *Chem. Eng. Sci.* **2013**, *115*, 88–94.

- (21) Kiedorf, G.; Hoang, D.; Müller, A.; Jörke, A.; Markert, J.; Arellano-Garcia, H.; Seidel-Morgenstern, A.; Hamel, C. Kinetics of 1-dodecene hydroformylation in a thermomorphic solvent system using a rhodium-biphospho catalyst. *Chem. Eng. Sci.* **2014**, *115*, 31–48.

- (22) Hentschel, B.; Peschel, A.; Freund, H.; Sundmacher, K. Simultaneous design of the optimal reaction and process concept for multiphase systems. *Chem. Eng. Sci.* **2014**, *115*, 69–87.

- (23) Brunsch, Y. Temperaturgesteuertes Katalysatorrecycling für die homogen katalysierte Hydroformylierung langkettiger Alkene. Ph.D. Thesis, Technische Universität Dortmund, Germany, 2013.

- (24) Schäfer, E.; Brunsch, Y.; Sadowski, G.; Behr, A. Hydroformylation of 1-dodecene in the Thermomorphic Solvent System Dimethylformamide/Decane. *Ind. Eng. Chem. Res.* **2012**, *51*, 10296–10306.

(25) Weidlich, U.; Gmehling, J. A Modified UNIFAC Model. 1. Prediction of VLE,  $h^E$ , and  $\gamma^\infty$ . *Ind. Eng. Chem. Res.* **1987**, *26*, 1372–1381.

(26) Ye, K. Process design based on CO<sub>2</sub>-expanded liquids as solvents. Ph.D. Thesis, Otto-von-Guericke-Universität Magdeburg, Germany, 2013.

(27) Steyer, F.; Flockerzi, D.; Sundmacher, K. Equilibrium and rate-based approaches to liquid-liquid phase splitting calculations. *Comput. Chem. Eng.* **2005**, *30*, 277–284.

(28) Yaws, C. *Chemical Properties Handbook: Physical Thermodynamics, Environmental Transport, Safety & Health Related Properties for Organic & Inorganic Chemicals*, 1st ed.; McGraw-Hill Professional: New York, 1998.

(29) Guthrie, K. M. *Data and Techniques for Preliminary Capital Cost Estimating*; McGraw-Hill: New York, 1969.

(30) Biegler, L. T.; Grossmann, I. E.; Westerberg, A. W. *Systematic Methods of Chemical Design Process*; Prentice Hall PTR: Upper Saddle River, NJ, 1997.