CHE 451, Chemical Engineering Design II, is described in the NCSU course catalog as:

“Chemical process design and optimization. The interplay of economic and technical factors in process development, site selection, project design, and production management. Comprehensive design problems.”

“Comprehensive” is an understatement. The course challenges students with a wide variety of chemical engineering processes, ranging from biotechnology to energy to fibers. In addition, students complement their technical skills with teaming, leadership, project management, and communication skills. Both the instructors and the students feel an enormous sense of pride as we view the accomplishments of almost seven months’ work.

We wish to thank those individuals and companies who have provided financial support or sponsored and advised projects this semester. The task of supporting and coaching these diverse projects and teams could not have been accomplished without their help!

Dr. Med Byrd (NCSU Department of Paper Science and Engineering)
Dr. Mari Chinn (NCSU Department of Biological and Agricultural Engineering)
Fluor Corporation
Dr. Russell Gorga (NCSU Department of Textile Engineering)
Dr. Alex Hobbs (NC Solar Center)
Dr. Henry Lamb (NCSU Department of Chemical and Biomolecular Engineering)
Dr. Thomas Losordo (NCSU Biological and Agricultural Engineering)
Mr. Michael Lowder (Eastman Chemical Company)
Dr. Jim McClain (Micell Technologies)
Mr. Rick Lawless (NCSU BTEC)
Dr. Behnam Pourdeyhimi (NCSU College of Textiles)
Mr. John Shell (Morris and Associates)
Mr. Al Springer (Springer and Associates)
Mr. Marcelo Tellez (NCSU BTEC)
Dr. Mark Van Dyke (Wake Forest Institute for Regenerative Medicine)
Dr. William E. Willis Jr.

Dr. Steven Peretti

Dr. Lisa Bullard
Two-Minute Project Overviews  
9– 9:45AM, Room 1025, Engineering Building II

<table>
<thead>
<tr>
<th>Board #</th>
<th>Title</th>
<th>Presenter</th>
</tr>
</thead>
<tbody>
<tr>
<td>1</td>
<td>Full Scale rhGH Production: Process Design and Considerations Utilizing the CHO Expression System</td>
<td>Ravi Appalabhotla</td>
</tr>
<tr>
<td>2</td>
<td>Production of Monoclonal Antibodies from CHO Cells in BTEC</td>
<td>Jason O’Neal</td>
</tr>
<tr>
<td>3</td>
<td>Fed-Batch Fermentation for eGFP Production in E. coli</td>
<td>Jake Davis</td>
</tr>
<tr>
<td>4</td>
<td>Cell Recovery Scale-Up</td>
<td>Steven Gregory</td>
</tr>
<tr>
<td>5</td>
<td>Chromatography Design</td>
<td>Wes Overton</td>
</tr>
<tr>
<td>6</td>
<td>Particle Formation of Invirase Utilizing Rapid Expansion of Supercritical Solutions</td>
<td>Mohamed Seyam</td>
</tr>
<tr>
<td>7</td>
<td>A Pilot Process for Production of Keratin Biomaterials</td>
<td>Ligaya Roque</td>
</tr>
<tr>
<td>8</td>
<td>Electrospinning Scale-up</td>
<td>Troy Gould</td>
</tr>
<tr>
<td>9</td>
<td>Hydroentangling Energy Reduction</td>
<td>Beth Duncan</td>
</tr>
<tr>
<td>10</td>
<td>Hydrogen Production as an Alternative Energy Resource from Coal Gasification Technology</td>
<td>Carl Denard</td>
</tr>
<tr>
<td>11</td>
<td>Coal to Synthetic Natural Gas</td>
<td>Michael Vergamini</td>
</tr>
<tr>
<td>12</td>
<td>Biomass Gasification</td>
<td>Jerome Savage</td>
</tr>
<tr>
<td>13</td>
<td>Bio-Mass Derived Synthesis Gas Cleanup for Fermentation to Produce Ethanol</td>
<td>Andrew LaGrange</td>
</tr>
<tr>
<td>14</td>
<td>Design of an Ethanol Pilot Plant using Wood Chips as a Raw Material</td>
<td>Katie Kennedy</td>
</tr>
<tr>
<td>15</td>
<td>Sweet Potatoes to Ethanol: Upstream</td>
<td>Kathy Fraley</td>
</tr>
<tr>
<td>16</td>
<td>Sweet Potatoes to Ethanol: Downstream</td>
<td>Renee Nobles</td>
</tr>
<tr>
<td>17</td>
<td>Biodiesel Pilot Facility Control System Design and Operation</td>
<td>Justin Harris</td>
</tr>
<tr>
<td>18</td>
<td>Design of a One Million Gallon per Year Enzymatic Biodiesel Production Plant</td>
<td>Robert Bullock</td>
</tr>
<tr>
<td>19</td>
<td>Algae to Biodiesel Conversion and Scale-Up</td>
<td>Samia Ilias</td>
</tr>
</tbody>
</table>

10 – 11:15AM  
Poster Session (First and Second Floor Atrium of Engineering Building I)

11:30 – Noon  
Closing Remarks and Special Recognition (Room 1025, Engineering Building II)
Since its FDA approval in 1985, recombinant growth hormone (rhGH) has shown to be effective in the treatment of both pediatric and adult growth hormone deficiency syndrome, Turner Syndrome, and chronic renal insufficiency. Recently, due to its anti-aging capabilities, it has even emerged as a popular product in the cosmetic industry. Since the market for rhGH is steadily increasing, the Help Barry Bonds Group (HBBG) is developing a facility for the commercial scale production of rhGH.

Market analysis suggests that HBBG could capture up to 45% of the market share by producing 20 kg per year of rhGH. HBBG is developing a novel, efficient process using Chinese Hamster Ovary (CHO) cells because of their ability to perform post-translational modifications and their ability to secrete proteins into solution. Design considerations at this stage will focus on fermentation and protein purification with the assumption that other upstream and downstream processes are previously established. The team has completed this stage of design including the steps below:

- Identified cell line
- Decided primary reactor design
- Estimated daily rhGH production
- Estimated rhGH yield/unit volume per 10 day run
- Identified purification unit operations
- Scaled all unit operations
- Developed potential purification design
- Determined cycle times for unit operations
- Identified key materials and costs
- Constructed Major Equipment List
- Constructed process flow diagrams for all upstream and downstream process
- Calculated Material Usage
- Estimated Capital Cost $200 M
- Estimated Utilities Cost
- Estimated Annual Material Costs
- Estimated Labor Costs
- Performed Economic Analysis

Efforts were focused on determining major costs, capital and operational, in order to determine whether this endeavor is a sound investment of company resources and if additional resource should be utilized to perform detailed engineering assessment as well as additional process development. Rate of return on investment was determined to range from 35% to 112% after 10 years.

After considering these results, the team believes that this would be a sound investment of company resources to advance to the next step of a detailed design and assessment of required resource and estimated profit. The team would also like to point out the rates of return on investment include paying 40% of profits to the owner to the expression system technology as well and the developers of the purification strategy. The team recommends that the company invests in developing similar technologies or methods that would be company owned or that the company purchases the actual technology or a similar in order to increase profitability.
Production of Monoclonal Antibodies from CHO Cells in BTEC

Melissa Bebb, Diana Bisbee, David Caudle, Dan Chamblee, Mark Hempel, Anthony Howard, Jason O’Neal, Lisa Saxon
Advisors: Dr. Henry Lamb and Mr. Rick Lawless

Growing demand for large scale production of new and improved bio-pharmaceutical products has provided North Carolina with an opportunity to become a major global biotechnology center. A large employment window for the surrounding economy is being created and North Carolina State University is working to meet this demand through the development of the Golden Leaf Biomanufacturing Training and Education Center (BTEC). BTEC has challenged our group to develop a teachable process and schedule for the production of monoclonal antibodies from Chinese Hamster Ovary (CHO) cells. This process will be used to give future students an understanding of real world industrial processes in the biopharmaceutical industry.

The upstream and downstream processes were designed to include state-of-the-art unit operations in order to maximize the experience that students will obtain with different equipment. These variations include a disposable 25L working volume bioreactor, an introduction of a microfiltration skid as a replacement to the traditional centrifugation system, and the use of membrane chromatography. Process parameters, pricing information and a commissioning strategy have been developed for the disposable bioreactor and membrane chromatography unit as requested by the client. The capital cost of additional equipment is approximately $225K. The addition of these operations will add the benefit of allowing students to learn using state-of-the-art equipment, without negatively affecting operating time or product yields. Existing equipment was determined prior to obtaining budgetary estimates for any additional equipment proposed for this process.

Equipment was sized and operating conditions were optimized to achieve the desired production of 200-400 \( \mu \)g of IgA per mL of media volume. The variable operating cost for the production per batch of IgA is estimated at $15K, which yields an annual production cost of $135K.

The production schedule was developed to ensure that a batch is produced once a month for nine months of the year. Total fermentation time from the initial inoculum of the 250 mL flask to the production bioreactor for batch and fed-batch processes were determined to be 288 and 408 hours, respectively. The total minimum downstream process time for both processes was determined to be 7.84 hours with an overall product yield of 88%. The batch process yields a total protein amount of 38.6 g, while the fed-batch process yields 77.2 g.

The development of the process for producing monoclonal antibodies from CHO cells will ultimately benefit all parties involved. Students will gain hands on experience with both commonly used and state-of-the-art techniques. This will increase the students’ curriculum experience and opportunity for better jobs. Companies will be able to hire highly trained employees directly from college and spend less money for on-site training. Finally, NC State will be recognized as a university that is at the forefront of the biotechnology advancement.
Fed-Batch Fermentation for eGFP Production in \textit{E. coli}

Andria Armstrong, Jake Davis, Shannadora Hollis, Brittany Lanier  
Advisor: Dr. Steven Peretti

North Carolina State University’s Biomanufacturing Training and Education Center (BTEC) will begin to offer courses in large-scale biomanufacturing. As part of the course, students will produce enhanced Green Fluorescent Protein (eGFP) by batch fermentation of recombinant \textit{E. coli}. However, the problem with batch fermentations is that lower yields of protein are obtained due to by-product formation. In the fermentation of \textit{E. coli} growing on glucose, acetate is formed as the by-product, and this leads to lower \textit{E. coli} cell densities and ultimately low eGFP production. It is known that fed-batch fermentation produces higher cell densities and thus increased protein production. The formation of acetate is minimized when glucose concentrations are low. By controlling the feedrate of glucose throughout fermentation, fed-batch fermentation allows \textit{E. coli} to grow to an optimum cell density, while limiting acetate formation. Therefore, the purpose of the current project was to develop a fed-batch fermentation process that limits glucose and results in higher cell densities and protein yield.

A method to control glucose levels is required for the optimization of the fermentation process. An automated glucose feeding strategy that avoids acetate accumulation is proposed. Using feedback control of the glucose feed rate through a peristaltic pump connected to the fermenter and associated control software, acetate accumulation is avoided and stable glucose concentrations are maintained.

The group met all of the specifications provided by BTEC. The fed-batch fermentation was designed according to the harvest goals of 20 g cells/L and 1.8 g eGFP/L. It was determined that fed-batch fermentation begins in batch phase and is followed by continuous phase, characterized by the start of glucose feeding. Following the batch phase, glucose feeding increases according to a quasi-steady state feed profile to facilitate optimal cell growth. Following the continuous phase, fermentation enters a final stationary phase, where the glucose feed is held constant at 0.2 L/hr. The total fermentation time was calculated to be 16.3 hours. At the end of fermentation, 384.57g eGFP are generated. The harvest goals were achieved from calculating material balances on the fermenter and downstream recovery and purification processes.

In order to accommodate the fermenter output and transition the current batch process to fed-batch, the HIC column and the UF/DF units needed resizing. Based on specifications, the HIC column control volume was calculated to be 34.01 L and the new height was determined to be 33.42 cm. The UF/DF filtration membrane was calculated to be 1 ft$^2$. Additional equipment needed includes a glucose feed container, a peristaltic pump to feed the glucose to the fermenter, and the associated tubing. It is also recommended that the HIC column and UF/DF membranes be purchased as well. New equipment prices for the HIC column, UF/DF unit, peristaltic pump, and glucose container are $20,038.00, $541.00, $1,500.00, and $140.00, respectively.

The critical process parameters such as eGFP concentrations, initial glucose concentration, glucose feedrates, the production rate, and equipment sizes have been determined. Using the data from the material balances done around the process and the teaching goals of running two fermentations per week, a production schedule has been done for each training week. These results were used to do an economic analysis of the process. Based on this analysis it was determined that the conversion from a batch fermentation process to a fed-batch process led to an increase in cost of approximately $1,480 per fermentation run.
Rapid growth of the biomanufacturing industry in North Carolina has created a surge in the demand for a properly trained workforce to operate these processes. In order to meet this demand, the GFP production process in the NCSU Biomanufacturing Training and Education Center (BTEC) will be used to train students in large-scale biomanufacturing practices. Since August 2006, Jellyfish Research has been designing Cell Recovery operations for this process which will involve both a disc-stack centrifuge, as well as a tangential flow filtration (TFF) apparatus.

The overall process is expected to use both TFF and centrifugation as follows. The first cell recovery step will take 250 L of fermentation broth from fermentation and will use a regenerated cellulose membrane to remove 200 L of broth in 3 hours using 0.5 m² of membrane material. The remaining 50 L of cells will be diluted back to 250 L using Sodium Phosphate and sent through a high pressure homogenizer to break open the cell walls. The second cell recovery step will use a disc-stack centrifuge to remove solid material from the product stream, leaving GFP and other proteins suspended in the liquid stream coming out of the centrifuge. This step is expected to be completed in 8 minutes. Chromatography and ultrafiltration/diafiltration will then be used to purify GFP from the other proteins.

With the system design in place, sizing of equipment and other construction materials were found. Then, protocols were designed for the safe operation of the experiments along with cleaning regimes. The design of the process equipment was then put into a simulation program to determine base flow streams and costs of equipment and operation. In addition, material balances were performed which agreed very closely to the simulation. The capital cost for the project was found to be $189,151 which includes installation, equipment, and basic piping. The operation cost was found to be $16,200 per year which includes cost of cleaning, operation, and replacement filters.

All objectives were completed and no additional time will be required. A recommendation for the process is after the construction is completed; practice experiments should be run. The practice experiments are to find any problems and optimize the experiment for easier operation. In conclusion, the designed cell recovery system will provide a good teaching method as part of the overall GFP production. Therefore, students of biomanufacturing should be well prepared for industrial careers after becoming familiar with operating this system.
As North Carolina continues to attract large and small biomanufacturing plants, a strong and skilled workforce becomes essential to operation. North Carolina State University and the Golden Leaf Biomanufacturing Training and Education Center (BTEC) are in the process of designing and implementing a pilot-scale processing plant to educate and train specialized workers for the emerging biotechnology industry. As part of the design process for the pilot plant, BTEC has requested that the Chromatography Senior Design Group efficiently design a chromatography system that would aid in the purification of green fluorescent protein (GFP) from recombinant *Escherichia coli* cells.

The goal of the project was to determine the following: column dimensions, type of chromatography, type of resin, operating conditions, and operating strategy. The operating conditions would include compositions, flow rates, and volumes of needed buffers for binding, washing, eluting, cleaning, and equilibrating. The optimal system would purify the GFP in the cell homogenate to a relative purity above 90% with an overall recovery of 85%. In addition to designing the chromatographic parameters, an economic analysis of the process was also requested.

In order to design an optimal chromatography system, the Chromatography group evaluated three different types of resin: 1.0 mL Q Sepharose FF Anion Exchange Chromatography (AEC), 0.6 mL Butyl Sepharose Hydrophobic Interaction Chromatography (HIC), and 0.4 mL Nickel Immobilized Metal Affinity Chromatography (IMAC). The resins were compared based on their binding capacity, elution strategy, and ultimately their purification of GFP. The binding capacity results showed that the IMAC resin was unable to effectively bind GFP, while AEC bound 1.219 mg of GFP and HIC bound 0.255 mg of GFP. A two-step elution strategy was developed for both the AEC and HIC columns. The GFP was eluted with 0.1M NaCl in the AEC column and with 1.25 M ammonium sulfate in the HIC column. There was no IMAC elution strategy performed due to the poor binding capacity of the resin with GFP. Lastly, the purity of the GFP procured from the developed elution strategy, was determined through densitometry. The samples containing GFP from the AEC column were an average of 95% pure as compared to average purities of 40.2% and 8.75% for GFP samples from the HIC and IMAC columns respectively.

The bench scale process was then scaled-up using Super-Pro modeling software, and an overall process flow diagram was simulated in Super-Pro to obtain the necessary material balances and economic analysis of the process. Calculations of the economic considerations led to an estimated cost of $14 million for the pilot plant.

After evaluating the indicated parameters and comparing the desired values provided by BTEC, the Chromatography Group strongly recommends the use of the Q Sepharose FF Anion Exchange resin in the pilot processing plant chromatography system. This resin exceeded the criteria purposed by BTEC and was cost effective when compared to the other resin choices.
Particle Formation of Invirase Utilizing Rapid Expansion of Supercritical Solutions

Allison Gibson, Chris Lowe, Joel Ortiz, Mohamed Seyam
Advisor: Dr. Jim McClain

The goal of the project is to design a semi-works scale of a supercritical process that has the purpose of decreasing the particle size of a drug to increase its bioavailability and therapeutic effectiveness. For this project, Invirase, a Human Immunodeficiency Virus (HIV) protease inhibitor, was chosen through research on market, size and production, and patent criteria. One of the problems with Invirase is that it has a low bioavailability (4%), meaning that the drug is not completely effective when it enters the body. In an attempt to increase the bioavailability, Roche, the producers of Invirase, decreased the particle size of Invirase.

Currently, the production of Invirase has been limited to microparticles that are processed in the form of a pill that is administered orally. This project will create a process that will allow for the production of nanoparticles of Invirase. The decrease in size of the particles, from micro- to nanoparticles, will allow for increased bioavailability, therefore increasing the amount of drug the body will actually receive per administered dose. The project explores the current technology of the Rapid Expansion of Supercritical Solutions (RESS) to produce nanoparticles and includes a process flow diagram that defines and visualizes the process. This project provides the specifications necessary to make the target amount of Invirase nanoparticles. A cost analysis has been performed to measure the competitiveness of the process in the market.

For the process to work, isobutylene is required as the solvent for Invirase in this RESS process. By using the process to decrease particle size, the surface area of the drug increases, thereby increasing the dissolution rate and bioavailability. The current process requires an annual production of 960 kg/yr of Invirase, with a target concentration of $10^{-4}$ mol Invirase/mol isobutylene ($1.20\times10^{-4}$g Invirase/g isobutylene). A recycle stream is included for economic factors associated with using isobutylene as the solvent.

Overall, the process has been deemed not feasible due to the Invirase’s properties, but will make a profit. Because of the high percentage of Invirase that is recovered in waste after being injected intravenously, the bioavailability of the drug will not increase significantly by changing the particle size alone. However, the process will make a yearly profit of about $23,000,000, making it greater than the current formulation ($22,000,000), making it economically advantageous assuming that the formulation produced by this project will be worth more than the current formulation. Even though the process may not work for Invirase, it can still work for other applications. If an application requires a hazardous solvent, the process designed for this project could be implemented.
A Pilot Process for Production of Keratin Biomaterials

Steven Aho, Chris Broda, Nargess Golafra, Erin Irons, Ligaya Roque
Advisor: Dr. Mark Van Dyke

Human hair keratins (HHKs) are the major components that comprise the make-up of human hair. Research into the potential uses of keratin began in the 1500s; however, scientists have only recently begun to realize the full medicinal benefits of these proteins. Keratin and its derivatives have the rare quality of being naturally biocompatible, making them great candidates for uses as therapeutics. Current research at the Wake Forest Institute of Regenerative Medicine (WFIRM) is generating groundbreaking results with HHKs. Preliminary studies have shown keratin to be promising in many areas including those of nerve regeneration, hemostatic and resuscitation agents, and wound healing accelerants. Continued research into the uses of HHKs may revolutionize many fields of medical treatment. One keratin derivative that shows exceptional promise in the wound healing arena is keratose.

Unfortunately, there is a lack of supply of keratose available to support the research demand, potentially hindering the future production of undiscovered keratose-based formulations. Research efforts at the WFIRM are impeded because the current process yields only 10-15 grams of keratin per batch, an amount inadequate to sustain the current research. A 100-fold scale-up pilot plant of the current keratin extraction process would provide a large facility capable of supporting in-house research, as well as generating additional revenue as a provider of keratin to other research groups and/or small commercial niches.

A joint venture between the North Carolina State University department of Chemical and Biomolecular Engineering and the WFIRM was formed to design this scaled-up pilot plant. The aim of the scale-up was specifically targeted toward keratose production; however, the facility was also designed to allow for the production of other keratin derivatives.

The process of scaling up keratose production involved the complete readjustment and improvement of the existing process. The basic protocol and reaction parameters were left unchanged; however, the pilot plant required entirely new equipment. Evaluations were performed on all technologies and appropriate selections were made. A novel stirring-mechanism utilizing magnetic technology was chosen for the reactor. To handle increased production, a continuous centrifugation unit and a tangential flow filtration system were employed. State-of-the-art thin-film evaporation and lyophilization technologies were selected for product processing. Simulations in SuperPro Designer helped to confirm the feasibility of the scale-up and the economic viability of the project.

Estimated equipment costs for the process totaled $638,500 with an additional $38,000 required for peripheral equipment. With the additional expenses associated with installation, piping, raw materials, utilities, and operator salaries, the total calculated cost of the biomaterial pilot plant was $1,112,000. Utilizing the facility as a purely commercial venture and marketing the product at half the price of similar, non-keratin-based wound-healing accelerants currently on the market showed the potential for annual revenues exceeding $8 million. Based on these potential profits and the current research demand for keratin, the following proposed pilot plant is worthy of further investment and development.
The expressed purpose of this project is to develop a scaled-up electrospinning device that will produce grams of electrospun polylactic acid fibers per day. The main reasons for scaling up the device include possible filtration and medical applications. The large surface area and porosity of the fiber mats produced may be able to trap submicron particles due to their permeability and selectivity. In the medical sector, these mats could be used in tissue scaffolding or drug delivery. Tissue scaffolding is a viable option if using an organic polymer, as it may be able to recreate the environment surrounding a wound in order to encourage cell growth towards the fibrous mat. Drug delivery may also be more efficient using a nanofiber mat because of the ability to dope the polymer with an antibiotic, which may maintain a more antiseptic environment during healing. The current production rate of nanofiber mats still remains on the order of micrograms, which is too low for any of these practical uses.

The development of prototype concepts was limited to novel approaches, with the focus on electrospinning from a thin film of polymer solution on a charged surface. A design concept was selected which pumps a polymeric fluid, 4% (wt/wt) polyethylene oxide for our tests, from a reservoir at the top of a twelve by six inch plate. In the initial design, a plate was charged using an electrode and high-voltage power supply. When solution flowed down the plate, the electrospinning jets traveled to a grounded collection plate roughly parallel to the falling film. The finer points of the design concept were considered as a Solidworks model was developed. After testing a crude design of the prototype, final decisions were made and an apparatus was constructed. The three main improvements made during the design process were the ability to control the flow plate angle, collection distance, and film flow as either streams or in a thin film. The apparatus consists of two pieces: one piece contains the flow plate and solution while the other piece holds the collection plate.

With a completed apparatus, tests were performed to determine the optimal conditions in which to create the electrospun nanofibers. Due to multiple variables, each one was tested individually to create fibers with the highest yield, smallest diameter, and most uniform morphology. The plate angle, collection distance, voltage, and flow type were varied and SEM images of the electrospun fibers were examined to determine what set of parameters created the best fibers. The prototype design successfully produced as high as roughly 0.4 g/hr of dry nanofibers with fiber diameters below 200 nm. This output represents roughly a 10-fold increase in nanofiber output compared to traditional syringe-type electrospinning processes. The flow rates and fiber morphology achieved suggest that the prototype concept may be appropriate for the production of nanofiber mats for commercial sale. With similar costs to traditional syringe-type methods and a substantial revenue advantage, the prototype design has the potential to become more profitable as industrial-scale nanofiber production becomes more prevalent.

The report outlines electrospinning design concepts, the theory behind their function, and the process and solution parameters that affect the process. Applications of this information to develop a scaled-up electrospinning design concept are also discussed. Lastly, the process of concept selection and testing, results of the design, and the economic and environmental implications are discussed in greater detail, along with recommendations for future work in electrospinning.
**Hydroentangling Energy Reduction**

Beth Duncan, Melissa Horton, Thomas Sinodis, Peter Stout, John Thompson  
Advisor: Dr. Behnam Pourdeyhimi

Hydroentangling is one of the most effective ways of making strong, durable nonwoven fabrics. The purpose of this project is to improve the hydroentangling process by developing a pre-splitting device, therefore reducing the number of passes required through the hydroentangler and the energy requirement. In this report, general information about the nonwoven industry, current methods of splitting bicomponent fibers, the existing hydroentangling process, and experimental results on mechanical and microwave splitting testing will be discussed. The research for this project will focus around the technology available at the Nonwovens Cooperative Research Center. Currently, the hydroentangling process at NCRC requires the fabric to pass through the system three times. This process involves a significant amount of time, energy, and equipment replacements, all of which result in a costly operation.

Throughout the last year, research has been done in several areas of nonwoven production to improve the current hydroentangling process. Mechanical splitting involves hydroentangling, carding, and needlepunching. Some problems associated with mechanical splitting are breaking of fibers, weakness of webs, and poor entangling for high performance fibers. Chemical splitting involves using a polymer that can be dissolved out of the fiber web. Some problems associated with chemical splitting are waste of raw materials, costs associated with clean-up, and environmental hazards.

Two methods of splitting fibers will be discussed, including a microwave system and mechanical shearing process. The microwave system involves bombarding a water-saturated fabric sample with microwaves in hopes to split the fiber components. With bicomponent fibers, it was proven that properties of both polymers can be utilized. In microwave separation, variation in the thermal expansion coefficients allowed for the polymers to expand at different rates, resulting in the bicomponent fiber splitting at the interface. Splitting was observed for the PE/PA6 polymer combination because of the large difference in the thermal expansion coefficients. The mechanical splitting system consists of a calendering roll that applies a high-pressure and/or shearing force to the fabric to induce fiber splitting. Results showed that the trilobal fiber formation is easy to split because the interface between the fibers is very accessible. It was proven also that fibers like polyethylene, which are softer in nature, are easily split with a lot of pressure. The two methods of pre-splitting are qualitatively analyzed.

A new process has been created for hydroentangling that requires about two thirds of the existing energy requirements and operates in a more environmentally friendly manner. Applying a high-pressured calendar roll causes the bicomponent fibers to begin to split. With some preliminary fiber splitting, the hydroentangler can more effectively entangle the fiber webs because it does not have to fibrillate the bicomponent fibers. This method has been proven to work with a smooth calender roll involving no shearing forces as well as a shearing calender with low pressure. The amount of fiber splitting before hydroentangling can be improved by using high pressure with a patterned calender and/or varying the speeds between the two calender rolls.
Coal to Hydrogen: A Study of Hydrogen Production as an Alternative Energy Resource from Coal Gasification Technology

Madeha Baqai, Carl Denard, Rahul Srinivas
Advisor: Mr. Mike Lowder

The energy requirement of the world population in commercial and industrial sectors is growing at a rate of 1.4% every year. Since coal is distributed widely across the U.S., it provides a feedstock resource that can be used to diversify the primary sources of energy. Hydrogen obtained from coal can become a resource for meeting domestic fuel and energy requirements, while its commercialization can minimize the emissions of greenhouse gases, decrease dependence on foreign oil, and provide a substitute to high natural gas prices. This design project focused on designing a hydrogen producing facility that uses 2500 tons/day of high volatile type A bituminous coal to produce 99.9% pure hydrogen since future customers, such as fuel cells and rocket fuel industries, require 99.9% pure hydrogen for operations.

The hydrogen facility design produces 100 million scf/day of hydrogen using a Texaco/GE Quench gasifier and a two-stage sour water-gas shift. The gasification cold gas efficiency was calculated to be 70% with 69.4 MBtu/lbmol HHV of hydrogen produced. The total capital cost of the facility is estimated to be $317 million for conventional gasification and $292 million using membrane-shift technologies. Gas-cleanup technologies include an activated carbon mercury removal, a Rectisol unit, a two-stage PSA unit, CO₂ sequestration and a Claus/SCOT unit. Manufacturing costs are estimated to be $119 million a year, requiring the hydrogen selling price to be $15.39/10⁶ Btu HHV based on 15% Rate of Return on Investment and 35% tax rate for 15 years.

The facility will be located in the southwest Alabama because of its proximity to the Gulf Coast and will be required to comply under Clean Air Interstate Rule and Clean Air Mercury rule issued by EPA. The facility design safety also requires all hydrogen transporting pipelines to be located above ground and away from all explosive or electric devices to avoid detonation or hydrogen-based fires. According to NASA regulations, packing hydrogen in while gray, ductile, or cast iron materials shall not be used in hydrogen safety.

Currently, 2007 natural gas price is 7.54/10⁶ Btu HHV yielding a hydrogen price of $10.56/10⁶ Btu HHV using steam reforming. Even though 80% of hydrogen today is obtained from steam reforming of natural gas, hydrogen obtained from coal gasification can become commercially competitive once depletion of natural gas reserves increases the price of energy. Therefore, further research in membrane technology can play an important role in producing hydrogen at lower costs since membrane technology can be used to directly purify hydrogen after gasification. Replacing the ASU with membrane separation technology to produce oxygen and using membrane shift reactors to directly purify hydrogen can also eliminate the need of a Rectisol and PSA unit. Understanding effects of salt oxides such as CaO technology to fix CO₂, and developing carbon sequestration techniques can improve syngas cleanup costs. Ultimately, a specific process configuration and alternative methods for gas cleanup designed for hydrogen production can reduce the price of hydrogen by 25% of the current cost with further R&D efforts.
Coal to Synthetic Natural Gas

Corey Allen, Lisa Downen, Stacy Fields, Chris Imbus, Mike Vergamini
Advisor: Mr. Mike Lowder

Currently, the United States consumes 43% more energy than it produces. As the prices of oil and natural gas continue to increase, coal is becoming more popular as an alternative energy source. This design project explores the conversion of high-sulfur coal, specifically Illinois No. 6, into pipeline-quality synthetic natural gas (SNG), which is primarily composed of methane. The conversion of coal to SNG typically involves two steps, the creation of syngas from coal and the conversion of syngas to methane. Each of these technologies has been studied and optimized by many organizations, including Eastman Chemical Company and the Great Plains Synfuels Plant; however, neither of these facilities would be effective for the specific purpose of converting Illinois No. 6 coal into SNG. The goal of this design project is to efficiently combine these technologies and to design a facility to profitably manufacture SNG with a large-scale commercial plant. The feasibility of this proposal relies on the price gap between coal and natural gas and the ability to convert coal to natural gas while minimizing energy loss.

Investigations into the technical aspects of the project began with studies of Eastman’s coal-to-methanol gasification process, and the Great Plains Synfuels Plant’s coal-to-synfuels process. From these investigations, the overall picture of a natural gas synthesis process was determined, and research was performed to compare and select the best individual process units specific for the project. These units were assembled along with theoretical process designs and tested through mathematical modeling with ASPEN simulation software.

This process consumes 3300 TPD of coal to produce SNG at a rate of 66.9 kmol CH₄/min, which equates to about 1x10⁷ MMBtu/yr. The yearly total expenditures of our plant, including capital, operational, and raw material costs, are estimated to be $63.9 million (real 2005 dollars). In order for this process to be profitable, this SNG must be sold at (Total Cost)/10⁷ dollars per MMBtu in real 2005 dollars. At the estimated plant cost, this is $6.39/MMBtu. The Energy Information Administration’s Annual Energy Outlook 2007 indicates the price of natural gas is expected to fall as low as $5.46/MMBtu (real 2005 dollars) in the year 2015, but then steadily increase to $6.52/MMBtu (real 2005 dollars) by the year 2030. Therefore, in the year 2030, this project is projected to be profitable. If elemental sulfur is sold as a byproduct, the SNG breakeven sale price becomes $6.22/MMBtu.

The current process design is not economically viable at the present time due to low natural gas prices and inefficiencies in the process. It is recommended to explore markets for byproducts, including elemental sulfur and carbon dioxide. Optimizing the heat exchangers in the process by creating a heat exchanger network and increasing the yield of individual methanation reactors by adding recycle loops will also offset design costs. Industries have had success in decreasing feedstock prices by negotiating with local and state governments. Discussions with state representatives for reduced costs for Illinois No. 6 coal are encouraged. Pilot plant experiments of methanation reactors can provide critical experimental data for a more accurate kinetic model. Furthermore, employing higher performing proprietary catalysts would permit greater efficiency and higher yield in the conversion to SNG.
Since August of 2007 the biomass gasification team has been planning, developing, and implementing a fully automated control system for the current bench-scale gasifier model. The model will be used to analyze the feasibility of large scale power production from gasification of biomass. The availability and price of waste biomass make this a potentially profitable venture.

Analysis of the gasifier model with respect to large-scale feasibility relies heavily on collecting large quantities of data. The current team has implemented an automated biomass feed and data logging system so that now the system can be adequately run by a single experienced operator or two inexperienced but technically competent operators. The automatic feed is controlled by temperature feedback based on maintaining a specific set point. The program has been set up such that changing this set point is only a matter of changing a single, specific number, making this setup highly efficient and effective.

The aforementioned setup required significant space; in particular the motor had to be mounted at the very top while keeping the whole unit sturdy. To mount and organize the devices a steel superstructure was designed and constructed. Not only did this facilitate implementation of the control system but also acted to greatly increase safety and security.

Finally, experiments were planned and completed which tested the automated setup’s effectiveness and determined essential operating parameters. The final experiment represents a culmination of this year’s design effort and was a great success in terms of automation and system integrity.
Bio-Mass Derived Synthesis Gas Cleanup for Fermentation to Produce Ethanol

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To sustain economic progress, alternative sources of energy will have to supplement finite petroleum supply. Considerable attention has been given to producing bio-ethanol from renewable sources such as corn, switchgrass, sugar cane and other biomass. One method of producing bio-ethanol converts synthesis gases (CO, CO$_2$ and H$_2$) generated by biomass gasification into ethanol and acetate by fermentation. The purity of the synthesis gas fed to the fermentation unit is critical because the *Clostridium ljungdahlii* bacteria that convert the gas components are inhibited by contaminants. Eliminating contaminants such as tars, chars, and ash improve the conversion rates producing more ethanol and acetate per unit of biomass. These tars can either be physically filtered out or chemically converted into useful substances. A combination of both is the best option, dedicating a physical separator to handle large particles while a catalyst handles the smaller contaminants.

For contaminant particles larger than 5 microns, physical removal was used. A 3” diameter cyclonic cleaner was designed and constructed to accelerate hot syngas in a vortex creating high-velocity cyclonic flow. Through centrifugal force, the gas and particles accelerate faster at the outside edge of the walls as it loops downward into the constricting conical section. Near the bottom of the cyclone, the gas makes a rapid change in direction flowing straight up to the outlet. Due to the contaminants higher density and thus higher momentum, they cannot make this change in direction and fall into the collection tube in the bottom. This design was established with the advice of Dr. Lingjuan Wang (BAE Department) and technical construction assistance from BAE machine shop.

The remaining contaminants can be reacted to form useful feed components through catalytic cracking. An alumina-doped zirconium oxide catalyst was chosen to replace the existing nickel and dolomite catalysts. This is an improvement because the zirconium catalyst is most effective at 550 °C, which is much lower than the 900°C required for comparable nickel and dolomite catalysts. A catalyst vessel with a heating element houses the zirconium catalyst to ensure that the catalyst does not fall below the ideal operating temperature of 550°C. This decision is backed by a research team led by S.J. Juutilainen in October 2006 that used a 36 cm$^3$ reactor bed of the zirconium catalyst to clean synthesis gas at a flow rate of 0.07cfm to achieve sufficient contaminant cracking. The temperature inside the catalytic vessel is measured, and the heating element can be used to add heat to the system if necessary.

Insulating the system is a method that can help maintain the thermal energy within the system. For a system with temperatures up to 1000°C like this one, ceramic insulation can be wrapped around the piping, cyclone and catalyst bed to reduce energy loss. This particular cleanup system is a small lab-scale system, and the hot gasification gas has very little residence time and loses very little energy prior to encountering the catalytic reactor.

There were delays in receiving syngas for cleanup, which was coming from another senior design group. Only one 30 minute cleanup run has taken place, from which successes and potential improvements can be gleaned. The cyclone filled with moist tars and operated at the expected pressure drop (5-6 inches H$_2$O) at 20 SCFM. The temperature of the syngas was measured as it entered the cleanup system and inside the catalytic reactor. The steady state inlet temperature was 353°C and the catalytic reactor temperature was 198°C. Calculating the heat loss through the system showed that the required inlet temperature for the catalytic reactor to operate at 550°C would be 1006°C. Had the inlet temperature been 1000°C as quoted, the constructed syngas cleanup system would be ideal.
Economic pressures from rising fuel prices have recently focused on fuel ethanol as an alternative energy source. Ethanol production rates in the United States have been increasing exponentially since 2002. However, for fuel ethanol production rates to continue to grow, alternative raw material sources will need to be investigated. Prospective raw materials include other forms of biomass, including wood chips, grasses, and agricultural residues. Currently, ethanol production from biomass is limited due to high processing costs. Therefore, it is beneficial to continue research into using biomass as a raw material for bioethanol production. This project focuses on the construction of an ethanol pilot plant research facility, under the supervision of the Department of Wood and Paper Science at North Carolina State University, to continue research into the conversion of biomass into fuel ethanol.

The production of ethanol from lignocellulosic materials consists of four main steps: pretreatment, cellulose hydrolysis, fermentation, and product purification. During pretreatment, the lignocellulosic materials are weakened, allowing for a higher conversion rate and separation into liquid and solid phase material. In the hydrolysis phase, cellulose chains will be decomposed into fermentable sugars, which are then converted into ethanol in the fermentation step. Separation, generally a form of distillation, allows for the purification of the ethanol product from aqueous solution. The proposed process for this pilot plant does not include a distillation section, as the volatility of ethanol introduces safety and explosion concerns. If distillation were used in this process, explosion-proofing of the entire lab would be required, causing costs to increase significantly.

The process to be used in the pilot plant for the conversion of ethanol has been selected. The raw material to be used is that of wood chips, and a final product of 99.5 wt% ethanol is the desired product. The pilot plant will be run in batches, using a feed of 200 kg of wood chips with a final product of about 15 gallons of 99.5 wt% pure ethanol. In the pilot plant, the separation phase of ethanol production will not occur due to safety regulations and the explosion factor of ethanol. Therefore, the ethanol produced will only be used for research purposes, with purity determined by analytical techniques. The pilot plant has been optimized to use not only equipment already present in the lab, but also to utilize the entire space set aside for the plant. This report details the chosen processes for the pilot plant. A process simulation was also created using SuperPro Designer. This not only allowed for a simulation of the system under specified conditions, but also allowed for the scaling and sizing of the major equipment needed for the plant. The total cost of equipment and installation is estimated to be on the order of 1.3 million dollars. The ethanol pilot plant will allow North Carolina State University to be a competitive research department in the field of bioethanol production.
Starch to Ethanol Team II designed the upstream portion of a process that will convert 10 lbs/hour of sweet potatoes to ethanol for North Carolina State University’s new Center for Integrated Biomass Processing. The upstream part of the process consists of processing sweet potato biomass and hydrolyzing it to 95% glucose for downstream fermentation. This pilot-scale plant will allow researchers to explore the feasibility of using sweet potatoes as an alternative to corn for ethanol production. If sweet potatoes are found to be a viable alternative to corn, it would mean a new and widely available feedstock to help meet the rising demand of ethanol in the United States. Ethanol and ethanol-gasoline blends are environmentally friendly, decrease dependence on foreign oil, and spur economic growth. This report will briefly outline the project, its outcomes, and recommendations for further process improvements.

The upstream portion of the sweet potato to ethanol process was designed as a batch process and modeled in SuperPro, a computer simulation program. In the simulation, the feed was defined as 65 lbs/batch of sweet potatoes (10 lbs/hour), which contains 32% starch, 66% water, and 2% impurities. For simulating purposes, starch was assumed to be a pure component of MW 162 g/mol that reacts with water (MW 18 g/mol) in a hydrolysis reaction to produce glucose (MW 180 g/mol). The hydrolysis reaction was catalyzed by the enzymes \( \alpha \)-amylase and glucoamylase. The addition of \( \alpha \)-amylase during the first stage of hydrolysis, called liquefaction, resulted in an overall 10% conversion of starch to glucose. Glucoamylase, added during the second stage of hydrolysis, called saccharification, resulted in an overall 95% glucose conversion or the production of 19.98 lbs of glucose from 20.8 lbs of starch in the feed. After simulating the process, a proposed layout of the 5000 square foot facility was constructed using AutoCAD, and the feasibility of designing the process to run continuously was studied. Additionally, relevant environmental regulations were evaluated and it was determined that the upstream batch process does not require any air permits. All emissions meet local, state, and federal environmental regulations. There are also no hazardous materials being used in the process.

An economic analysis, which included capital and manufacturing costs, was performed on the upstream process based on simulation results. The capital costs for the process are estimated to be $97,000. This cost excludes equipment for pretreatment processing of the potatoes. The NCSU Animal and Poultry Waste Management Center, located near the facility, owns equipment for this purpose and has agreed to process the potatoes at a cost of $575 for 500 pounds. An electric boiler for supplying steam for the heating processes also exists on site and is not included in the cost estimate.

The costs of enzymes, utilities, and pretreatment contribute to the manufacturing costs of the process. The initial manufacturing cost is estimated at $76.00 per batch. The current market price of ethanol, according to the Chicago Board of Trade, is $2.15 per gallon. This process is estimated to produce 2.84 gallons of ethanol per batch. At this rate, the process will lose $69.40 per batch.

This final report presents a batch process design of the upstream portion of a process that converts sweet potatoes into ethanol. The topics covered include the full motivation for the project, technical background, technical modeling results, process economics, and an evaluation of environmental concerns. Recommendations for future work are also included. The next stage in the design of the facility is to combine the upstream and downstream process designs and optimize the conversion of sweet potatoes to ethanol.
In recent years, the United States has recognized the need to develop fuel alternatives. The Department of Energy believes the use of biomass-based energy to be advantageous for national energy security, economic growth, and environmental sustainability. To help address this need, Sweet Energy has been designing an ethanol production facility utilizing sweet potato biomass since August of 2006. The facility will be capable of processing 10 pounds of sweet potato biomass per hour. This project supports the goals for North Carolina State University’s Center for Integrated Biomass Processes, which will focus on developing new technology for the production of fuel ethanol from different biomass resources found in North Carolina. The focus on Sweet Energy’s efforts has been designing the downstream portion of the pilot production plant, which includes fermentation and ethanol recovery.

The pilot plant has been simulated with SuperPro modeling software. The simulation model begins with an initial charge to the fermenter consisting of an intermediate stream and a culture composed of nutrients, yeast (*Saccharomyces cerevisiae*), and water. The solution is then heated, agitated, and allowed to ferment. After the stream leaves the fermenter, it passes through a centrifugation unit where the yeast and other biomass are removed. To reduce cost and increase reaction rates, yeast is recycled to the fermenter. A stream containing mostly ethanol and water is sent from the centrifugation unit to the ethanol recovery section where two distillation columns and a molecular sieve are utilized to achieve the desired ethanol concentration of 99.6% ethanol. The requirements set by the American Society for Testing Materials for "denatured fuel ethanol" are a minimum of 92.1%v/v ethanol and a maximum of 1%v/v water. Both of these standards are met by the simulated facility.

Based on the above simulation model, the major equipment for the plant includes a reactor, centrifugation unit, two distillation columns, a molecular sieve, and four pumps. The total cost of this equipment for the plant will be approximately $216,000. In addition to the equipment, the cost of total utilities for the plant was estimated at $74,800/yr. This figure includes steam, cooling water, and electricity needed to run the pilot plant, and is based on operating the plant forty hours a week for fifty weeks of the year.

In addition to designing the facility, major regulations for the facility were investigated. The major regulating agency pertaining to ethanol production facilities is the Environmental Protection Agency (EPA). Most of the legislation regulating ethanol production facilities pertains to the air emissions; however, emissions from the pilot plant are not expected to violate regulations.

This report presents the downstream portion of an ethanol production facility based on SuperPro simulations and technical literature reviewed. A detailed technical background and major findings are also presented in this report. Overall, the project team achieved the goal of designing the fermentation and product recovery sections of the facility.
Since September 2006, team BioEvil has been working to develop and implement a control system for a pilot scale biodiesel production plant. Construction of a pilot scale biodiesel production plant in the Waste Processing Facility at NC State’s Animal and Poultry Waste Management Center is now completed, and to date over 200 gallons of clean biodiesel have been produced. The purpose of this project is to reduce the labor required in the batch production of biodiesel, with the potential for the current biodiesel production plant to be converted to run as a continuous process if the demand for higher production rates should arise. The results of this project are a semi-automated biodiesel production plant, and this summary briefly outlines the final status of the project.

Through manual runs of the reactor system, a proper production procedure has been established. For this production procedure, flow diagrams were constructed. From these diagrams, points of interest were identified as optimal control variables for automation. More specifically, it was determined that the reactor mixing and settling phases should be automated, along with the wash water temperature, flow, and drain, followed by the final drying in the reactor. Also, temperatures should be monitored and recorded. Likewise, the feedstock addition should be gauged by load cells. To accomplish these goals, a Programmable Logic Controller (PLC) was necessary, along with all its related parts.

The PLC and related parts were programmed and wired to control the optimal automation variables listed above. Also, the process variables are adjustable through the visual PLC interface that has been programmed to accept operator input. To begin the process, the operator inputs the desired number of gallons of soy oil to be reacted into the interface. The PLC monitors two load cells during addition of the feedstocks and lets the operator know when the proper amounts are in the reactor. The operator then inputs desired reaction time, settling time, and temperature, and the PLC controls three stainless steel solenoid valves and thermocouples that fully control the reactor mixing and settling phases. After a manual glycerin drain, the operator then inputs the desired wash water volume and temperature, along with the settling time. The PLC then facilitates the transfer of the raw biodiesel to the wash tank, controls the wash water flow with an automated mixing water temperature control unit, and controls the drain time with another solenoid, allowing full automation of the wash phases. Finally the PLC facilitates the transfer of the washed biodiesel back to the reactor for a final drying, with the time and temperature input from the operator. With theses processes automated, the goal of two batches per day for 1000 gallons per week of biodiesel production is attainable.

The final cost of the project, not including feedstocks, electricity, and water costs was $5247.68. This cost includes $2504.53 for the PLC system and all its accessories, such as the interface, relays, and mounting hardware. It also includes $2037.28 for the control devices and sensors, such as the solenoids, load cells, and thermocouples. Finally, it includes $705.87 for lab equipment to remain at the facility for titrations and testing. This report includes justification for these purchases and full descriptions of all equipment purchased.

The design is currently set up to run on low free fatty acid (FFA) feedstocks, but the PLC was programmed in such a way to allow for a pretreatment step when the switch to high FFA feedstocks is desired. Likewise, future considerations for automated oil and methoxide addition are also possible with
the PLC, along with the glycerin drain after the reaction is complete. These considerations could further increase the efficiency and thus production capacity of the facility.

**Board Number: 18**

**Design of a One Million Gallon per Year Enzymatic Biodiesel Production Plant**

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Dependence upon foreign sources of oil, increasing concerns over petroleum fuel emissions, and the rising price of oil have recently led to an increased interest in alternative fuels in the United States. Biodiesel is one such alternative fuel that has shown promise as a possible replacement for petroleum diesel. Biodiesel fuel can be produced from existing feedstocks, such as plant oil or animal fats, and can be used in conventional diesel engines without modification. Furthermore, the emissions characteristics of biodiesel, including SO$_x$, particulate matter, unburned hydrocarbons, and life cycle CO$_2$ production, are environmentally friendly compared to the emissions from petroleum diesel.

This report summarizes the results of a project taking advantage of the increased demand for biodiesel by designing a biodiesel production plant located in North Carolina. This plant will produce 1.27 million gallons of biodiesel per year, utilizing a locally available animal fat feedstock from Fayetteville-based Valley Proteins and a novel whole cell enzymatic catalysis method. The project includes all phases of plant design, from feedstock selection and plant site selection to economic analysis and environmental regulations. The final goal of the project has been to create a computer simulation of the biodiesel production facility and use the simulation to compile a study estimate economic analysis of the proposed plant.

To begin, the design group completed a detailed literature review of the biodiesel production industry. An animal fat feedstock was identified in the form of a poultry fat source from Valley Proteins, and composition data from this feedstock was used to determine the physical properties of the feedstock. A plant site was identified near the feedstock source and along Interstate 95, allowing for simple transportation logistics. The plant unit operations, including the main reactor and much of the downstream purification and separation equipment, were based upon the literature review, in-class assignments, and conversations with the project advisor. Using this information, a detailed computer simulation has been created using SuperPro Designer to yield estimates of equipment sizes and costs as well as raw material requirements.

The final plant design recommendations call for annual raw material requirements of 4.2 million kg of poultry fat, 1.8 million kg of water, 935000 kg of methanol, as well as Ca-alginate and media for cell growth, all at an estimated annual cost of approximately $2.17 million. The primary product will be 99.9% pure biodiesel, which will be sold at $2.75 to be competitive with petroleum diesel and will net yearly revenues of $3.4 million. The glycerol byproduct of the biodiesel production will be purified to pharmaceutical grade (99.5% purity) and sold to yield an additional $447000 in revenues. Total capital costs are expected to be roughly $1.05 million for the plant equipment and $250,000 for the land. In total, after factoring in several tax breaks offered to North Carolina biodiesel producers, the plant is expected to have a payback period of 4.7 years and a return on investment of 21.24%.
Algae to Biodiesel Conversion and Scale-Up

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The goal of this project was to successfully produce one liter of biodiesel from *Scenedesmus* algae, a strain of algae common to North Carolina. In order to accomplish this goal, the project was divided into four distinct steps: growth, harvesting, extraction, and conversion. *Scenedesmus* algae have been growing in the Undergraduate Teaching Laboratory since December 2006 in a photobioreactor (PBR). Algae were periodically separated from water by harvesting algae through flocculation methods. A crude product was then extracted from the algal biomass using solvent extractions. A SuperPro simulation in addition to a pond system design was used to determine the economical feasibility of a large-scale plant to grow and convert algae to biodiesel.

Algae were grown in an indoor PBR. Algae concentration reached a maximum of approximately 0.05% algae (0.5 g dry weight/L water). The algal species *Scenedesmus* was chosen because it is native to North Carolina and exhibits a high natural oil content ranging from 14 to 40% by weight. In order to separate algae from the suspension, flocculation followed by either dissolved air flotation (DAF) or sedimentation was used. Chitosan and alum were both tested as possible flocculants, with alum providing a better separation of algae from the suspension. By using DAF, algae concentration was increased by two orders of magnitude. Sedimentation provided an average algae removal of 0.32 grams dry weight algae removed per liter, whereas the highest algae removal obtained by DAF was 0.18 grams dry weight algae removed per liter. While sedimentation provides for greater removal, it requires a longer process time.

Ethanol, hexane, and a Bligh and Dyer mixture consisting of chloroform and methanol were all used as solvents for oil extraction from algae. H-NMR analysis of the Bligh and Dyer product confirmed that oil was extracted from algae grown in the PBR. Approximately 20% of the algae dry weight was extracted as oil using this method. The ethanol extraction method also resulted in a product; however, further analysis was not conducted because there was not enough product to perform an NMR. The hexane extraction method did not result in any product.

Simulation results prove that an algae-to-biodiesel production facility capable of producing 1000 gallons of biodiesel per batch is not economically feasible. Algal biodiesel would have a breakeven selling price of $4.96, pricing the product out of the current biofuels market. The processing facility would have a total capital investment of $5.92 million with the contributions of the production plant and pond system costing $1.014 million and $4.906 million, respectively. In addition one production batch of biodiesel would require a single pond to span 10 acres capable of holding a volume of 38.5 million liters. This volume assumes that the concentration of algae in the pond water is approximately 0.05%. In order to control speciation and allow for sufficient algal growth, a system of 30 ponds harvested every three weeks would be required. This entire system would span over 300 acres.

The goal of producing one liter of algal oil was not achieved during the course of the project. The major limiting factor in extracting oil from algae was the concentration of the algae in the PBR. Because the DAF could only process 14 liters of algal suspension per batch, the low concentration of algae in the PBR prevented appreciable amounts of algae from being obtained. Therefore, larger equipment is required to remove greater amounts of algae using flotation. The low algae concentration is also a major factor in determining the feasibility of building an algae-to-biodiesel conversion facility. At the current concentration of 0.05%, the process is not economically feasible. Increasing this concentration could make this process feasible.